A REVIEW AND ASSESSMENT OF PROCESSES FOR INDIRECT LIQUEFACTION OF ONTARIO LIGNITE INTO METHANOL AND OTHER LIQUID FUELS - A 50/50 SHARED-COST CONTRACT

FINAL REPORT

BY

ONTARIO ENERGY CORPORATION

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INTRODUCTION

The Methanol Assessment Program was undertaken by the Ontario Energy Corporation (OEC) under contract to the Ontario Ministry of Energy. The Ministry commissioned the work as part of the Alternative Transportation Fuels Program which encourages the substitution of other fuels for petroleum products in the transportation sector.

One of the objectives of the Methanol Assessment Program is to identify the best technology and feedstock combination(s) for production of methanol in Ontario.

Lignite, wood and peat were selected as the main indigenous feedstocks. The lignite based options were studied under a separate program co-shared by OEC and CANMET as defined by contract OSQ82-00201 dated December 15, 1982.

A similar study, also sponsored by CANMET, was previously undertaken by SASKOIL for the Saskatchewan lignite. This study was prepared by SNC and was released in early 1983. It identified five (5) gasification technologies for the indirect liquefaction of Saskatchewan lignite to methanol:

High Temperature Winkler (HTW)
Koppers-Totzek (K-T)
Lurgi
Texaco
Westinghouse

Given the similarities which exist between Saskatchewan and Ontario ligite deposits, we preselected these five gasification technologies as a basis for our comparative techno-economic assessment.

The present study was carried out in the following sequence:

a) A technical evaluation of the preselected technologies, together with a capital and operating cost estimate was performed on various processing options in terms of natural gas and electrolytic hydrogen co-feedstocks, power co-generation and plant capacity. This evaluation lead to the selection of the HTW technology, for which gasification tests were conducted in Germany and cost estimates developed in more detail.

b) An overall economic analysis of the various processing options was then developed for the HTW technology. The straight lignite-to-methanol option (LTM), without any co-feedstock was identified as the most promising. Discounted cashflow returns on investment were developed and a preliminary financial analysis was included.

The technical evaluation was sub-contracted by OEC to Kilborn Ltd. of Toronto. This work was originally intended to be executed in four successive phases:

- Phase 0: to confirm the technical suitability of the preselected gasification technologies for the treatment of Ontario lignite.
- Phase 1: to select, on a preliminary basis, the most promising technology and the most attractive processing option.
- Phase 2: to coordinate with the selected technology developer a program of lignite gasification tests.
- Phase 3: to proceed with preliminary engineering and detailed cost estimate of a lignite to methanol facility and to evaluate the Mobil methanol-to-gasoline process (MTG).

During the execution of this work, the OEC and its Advisory Committee, composed of representatives of SASKOIL, ONAKAWANA DEVELOPMENTS LIMITED and CANMET, decided to bring some changes to the original scope of work.

- Phase 0 was carried out as planned
- Phase I was slightly expanded to include alternative options based on "pressure swing absorption" (PSA) and power co-generation.

- In Phase 2, the HTW gasifier was tested with both Saskatchewan and Ontario lignite samples. The gasifier's operating characteristics were extrapolated to a commercial scale plant and a more detailed material and energy balance was developed for the integrated methanol plant. This resulted in a more accurate capital and operating cost estimate. All this additional work was labelled "Phase 2A".

- Phase 3 was entirely cancelled. The additional work carried out in phase 2A was deemed sufficient, at this stage, to assess the lignite to methanol process. The evaluation of the MTG process was not considered necessary, since this option had already been explored, to some extent, in the SASKOIL/SNC report. It was also generally agreed that a lignite based methanol plant could only become viable after 1990, and therefore could not justify the preliminary engineering effort at this time.

This report is presented as follows:

- Chapter I summarizes the findings of the study in technical and economic terms. Recommendations for future action are also stated.

- Chapter 2 highlights the basic concepts of methanol production from carbonaceous materials and introduces the theoretical advantages which can be derived from the use of natural gas or electrolytic hydrogen as co-feedstocks.

- Chapter 3 outlines the method used for economic evaluation. The relatively large number of options considered calls for a sequential evaluation process which screens the options in terms of potential economic attractiveness and allows a more in-depth analysis of the most promising ones.

- Chapter 4 reviews the main features of the conventional natural gas to methanol process (GTM). The GTM process is viewed as the reference process against which any alternative must compete.

- Chapter 5 is a summary of the KILBORN study and the RHEINBRAUN report of lignite testing.

- Chapter 6 reviews the basis for all feedstock prices, i.e. lignite, power and natural gas.

- Chapter 7 presents a comparative economic evaluation of the three processing options based on the HTW technology, and the corresponding conventional process. These options are:
  - o use of lignite as the sole feedstock (LTM)
  - o use of natural gas co-feedstock (LGTM)
  - o use of electrolytic hydrogen co-feedstock (LHTM)
  - o use of the conventional gas to methanol process (GTM*)

- Chatper 8 deals with a discounted cash flow analysis of the most attractive option (LTM) using several sensitivity cases. The same analysis is performed on the GTM* option for comparison purposes.

- Chapter 9 presents a typical financial analysis of the LTM option to derive the possible return on equity.

The KILBORN study, the RHEINBRAUN test report and the computer printouts for economic analysis are assembled in the Appendix.

I – FINDINGS AND RECOMMENDATIONS

The assessment of alternative methanol processes leads to the following conclusions:

1.1 TIMING

Alternative processes, based on Ontario's lignite, are not likely to be competitive with the conventional GTM* process in the 1980's.

Furthermore, several factors can delay the implementation of such projects, beyond 1990:

1.1.1 Improvements of the GTM Process

Increasing natural gas prices, would provide greater incentives for low energy designs. This phenomenon is quite apparent in the ammonia industry where two high efficiency world scale plants are scheduled to come on stream shortly.[1] Such improved efficiency could translate into savings of 0.5 to 1.0¢/litre of methanol.

Chem-Systems Inc. (New York) has been developing a new converter in which synthesis gas is sparged through a light oil bath containing suspended catalyst granules. The improved heat transfer system allows a higher feed gas concentration resulting in 5 to 10% decrease in cost. A pilot plant is being built in LaPorte, Texas.

The use of an oxygen fed secondary reformer is being investigated. Air fed secondary reformers are commonly used in ammonia plants. If air is replaced by oxygen, natural gas can be reformed into hydrogen and carbon monoxide, in the proper stoichiometric ratio of 2:1. This would result in a higher overall efficiency.

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[1] The Sherrit Gordon Mines plant is already operating in Alberta. The CIL plant is scheduled for start-up in Ontario in 1985.

1.1.2 Higher Methanol Price

If methanol selling price, at a southern Ontario plant gate, rises significantly above 20¢/litre in 1990, in constant 1983 Canadian Dollars, the least capital intensive option is favoured. The conventional GTM* process would therefore be more attractive than any one of the more costly alternative routes.

1.1.3 Lower Natural Gas Price

If actual gas prices are below the low case projections established by the Ontario Ministry of Energy, the GTM* option would offer economic advantages difficult to compensate by lower lignite prices.

1.2 CO-FEEDSTOCK APPROACH

The rationale of an alternative methanol process rests on the premise of a steady escalation of natural gas prices. Therefore, alternative processes using substantial amounts of natural gas as a co-feedstock should not be envisaged.

The lignite-gas option (LGTM) uses natural gas for 50% of its total energy requirements and was not found competitive with the straight lignite to methanol (LTM) process.

While electrolytic hydrogen can enhance the economics of methanol processes based on peat or wood, it is not an option we would recommend for the lignite based process for the following reasons:

- There is no incentive in displacing low cost lignite ($1.30/GJ) by electric power priced at about 2¢/kWhr.
- The use of electrolytic hydrogen would reduce the lignite consumption by 40% and consequently would increase the cost of the mining operations, for a given plant capacity.

- The hydrogen concentration in the raw gas of the gasifier's outlet is relatively low, as compared to wood or peat gasification. Therefore the required electrolytic hydrogen plant should be significantly larger and more costly.

Neither co-feedstock is recommended for the gasification of lignite into methanol.

1.3 RECOMMENDED PROCESSING OPTIONS

The demand for fuel methanol dictates the plant size and, in turn, the selection of the processing option.

A lignite based large scale plant, producing one billion litres annually, (2500 TPD), could come on stream in the mid-1990's. This plant should be based on lignite alone. It was assumed to be built in the Onakawana area, but a more detailed site study could highlight advantages offered by a southern Ontario location.

This large scale plant generates an internal rate of return of 9.5%, in real terms, before taxes. The commercial viability of this project could be enhanced through leveraged financing and an appropriate fiscal structure.

1.4 SELECTION OF GASIFICATION TECHNOLOGY

The High Temperature Winkler (HTW) technology, developed by Rheinbraun-Uhde (West Germany) was found suitable for the gasification of lignite.

Among developing technologies, the HTW gasifier is almost commercially available. A large scale demonstration plant, operating on German lignite is scheduled to come on stream this year near Cologne.

Gasification is a field undergoing considerable research. More efficient designs may be offered in the not so distant future. The final selection of the gasifier should be made when the project is implemented.

2 — PROCESS HIGHLIGHTS

This Chapter highlights the technical steps for processing each type of feedstock, setting the stage for the technical and economic evaluation. It also summarizes the main factors that have to be considered in judging the technical performance, overall efficiency and economic viability of a process. The detailed technical analysis of processing options for each of the four feedstocks is contained in the appendix to this part of the report.

Methanol's chemical formula is CH_3OH. With only one carbon in combination with hydrogen and oxygen, it is the simplest alcohol. The formula can also be expressed as MeOH, where Me stands for the methyl radical CH_3.

Methanol manufacturing processes involve two main steps:

- Production of a **synthesis gas** containing primarily carbon monoxide (CO) and hydrogen (H_2);
- Catalytic conversion of this synthesis gas into methanol by the **exothermic reaction.**

$$CO + 2H_2 \longrightarrow CH_3OH \quad (\Delta H = -24.6\ \text{kcal})$$

Small amounts of carbon dioxide (CO_2), contained in the synthesis gas, are also converted into methanol.

$$CO_2 + 3H_2 \longrightarrow CH_3OH + H_2O$$

Our analysis for the second step -- catalytic conversion -- is based on the commonly used low pressure technology developed by Imperial Chemical Industries Limited (ICI). The method of producing synthesis gas in the first step depends on the feedstock.

Potential feedstocks differ in terms of the amount of carbon and hydrogen that they contain. For example, the most widely used feedstock, natural gas contains more hydrogen than is needed relative to its carbon content. The reverse is true for carbonaceous feedstocks such as lignite and wood.

The technical objective is to make the energy efficiency of the combined process as great as possible while minimizing the loss of valuable carbon or hydrogen. There are a number of possible processing options, including the use of co-feedstocks or combined feedstocks where one feedstock provides relatively more carbon and the other more hydrogen. But all options involve trade-offs and all have some disadvantages, even in the case of the well-developed and proven natural gas processes. Determining the best option for new feedstocks such as lignite, requires a comprehensive evaluation.

2.1 METHANOL FROM NATURAL GAS

About 80% of the world methanol production capacity uses natural gas as a feedstock.

Steam reforming of natural gas is a proven process used in both methanol and ammonia plants. This high pressure, high temperature, catalytic **endothermic reaction** leads to a hydrogen-rich synthesis gas.

$$CH_4 + H_2O \longrightarrow CO + 3H_2 \quad (\Delta H = +54.2 \text{ kcal})$$

The hydrogen to carbon monoxide ratio H_2/CO is 3:1 in the synthesis gas. The methanol synthesis requires a **stoichiometric ratio** of 2:1.

$$CO + 2H_2 \longrightarrow CH_3OH$$

Because the synthesis gas is hydrogen-rich, a significant portion of the valuable hydrogen gas must be purged from the synthesis loop and is used as fuel to provide part of the heat required by the steam reforming.

2.2 LIGNITE FEEDSTOCK

In theory, natural gas can be replaced by any carbon based solid feedstock. The gasification of such a feedstock leads to a hydrogen-poor synthesis gas.

$$C + H_2O \longrightarrow CO + H_2$$

Here, the H_2/CO ratio equals one and is not suitable for methanol production. But, this synthesis gas can be enriched in hydrogen through a CO **shift conversion** by steam oxidizing part of the carbon monoxide into carbon dioxide.

$$CO + H_2O \longrightarrow CO_2 + H_2$$

To get enough hydrogen, a large portion of the carbon monoxide has to be shift converted into CO_2 and H_2. Much of the CO_2 produced has to be vented into the atmosphere and is lost in the overall balance.

2.3 NATURAL GAS AS A CO-FEEDSTOCK

Synthesis gases obtained through gasification of carbonaceous feedstocks and through steam reforming of natural gas have complementary hydrogen content. The first is hydrogen-poor and the second, hydrogen-rich. This can be used to advantage in combined feedstock processes.

The hydrogen-poor gas, obtained by gasification of lignite, can be mixed with the hydrogen-rich gas from the steam reformer in proportions that ensure the final synthesis gas has the required H_2/CO ratio of 2.

Figure 2-1 illustrates this concept. In theory, combining the feedstocks should significantly increase the yields .

The hydrogen and carbon oxides produced by the gasification and steam reforming reactions can be fully used. However, a closer examination of both reactions points out some of the limitations of this system.

2.3.1 Synthesis Gas (Syngas) Composition

If CO and H_2 were the only components of the synthesis gas, methanol synthesis would occur as:

$$CO + 2H_2 \longrightarrow CH_3OH$$

But CO_2 is always present and some is required in the catalytic conversion. As the next equation shows, CO_2 absorbs 1.5 times as much hydrogen as does CO to make the same quantity of methanol.

$$CO_2 + 3H_2 \longrightarrow CH_3OH + H_2O$$

As a result, the components of the syngas have to be in these proportions:

$$H_2 = 2CO + 3CO_2$$

In other words, sufficient hydrogen must be available to react with both the CO and the CO_2. Instead of the oversimplified H_2/CO ratio of 2, therefore, the actual ratio R is:

$$R = (H_2 - CO_2)/(CO + CO_2) = 2$$

2.3.2 Gas From Steam Reforming

Steam reforming of natural gas primarily generates CO and H_2 with some CO_2. The ratio $(H_2 - CO_2)/(CO + CO_2)$ approaches 3.

2.3.3 Raw Gas From Gasifier

Depending on the operating conditions in the gasifier - temperature, pressure - the raw gas composition will vary.

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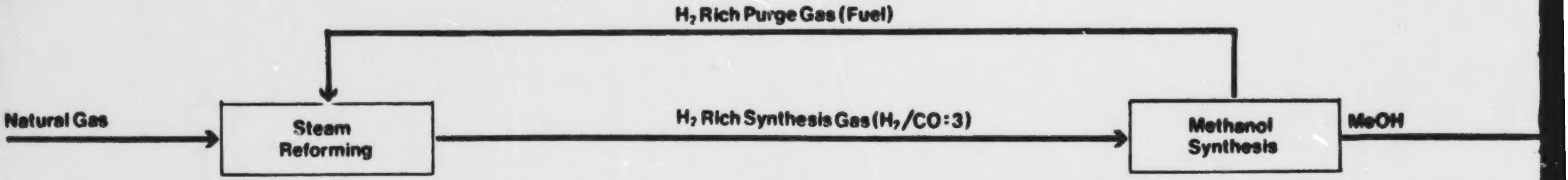
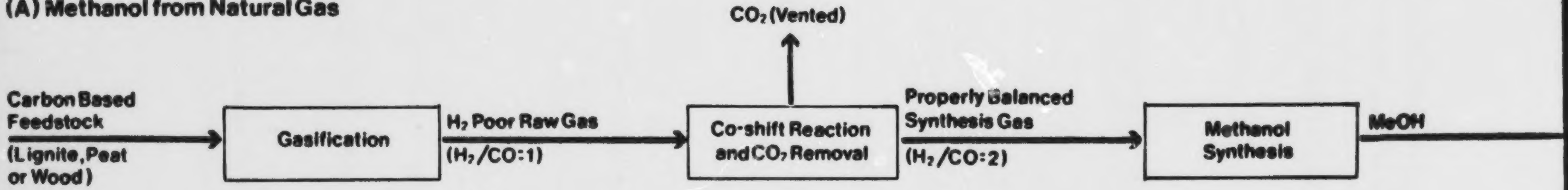
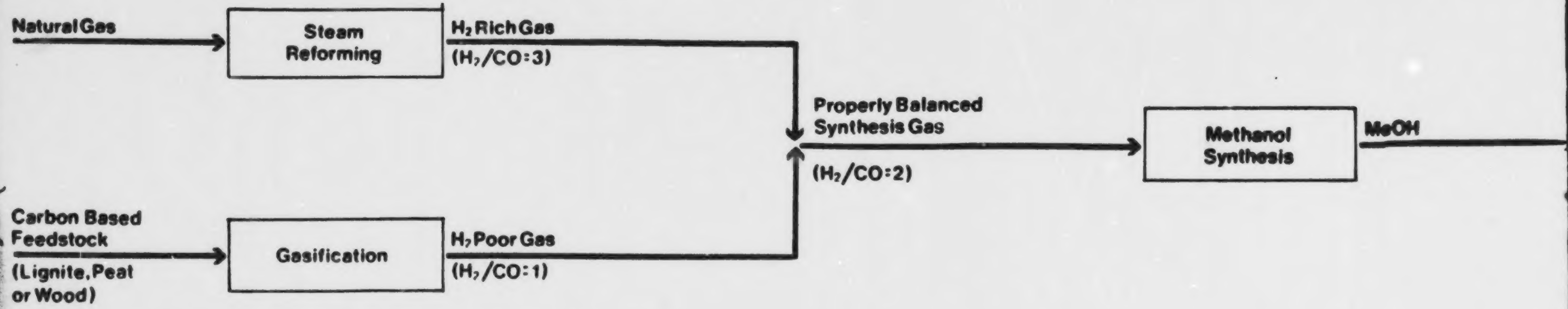

H₂ Rich Purge Gas (Fuel)
Natural Gas
Steam Reforming
H₂ Rich Synthesis Gas (H₂/CO:3)
Methanol Synthesis
MeOH

(A) Methanol from Natural Gas


CO₂ (Vented)
Carbon Based Feedstock
(Lignite, Peat or Wood)
Gasification
H₂ Poor Raw Gas
(H₂/CO:1)
Co-shift Reaction and CO₂ Removal
Properly Balanced Synthesis Gas
(H₂/CO:2)
Methanol Synthesis
MeOH

(B) Methanol from Peat, Lignite, or Wood


Natural Gas
Steam Reforming
H₂ Rich Gas
(H₂/CO:3)
Properly Balanced Synthesis Gas
(H₂/CO:2)
Methanol Synthesis
MeOH
Carbon Based Feedstock
(Lignite, Peat or Wood)
Gasification
H₂ Poor Gas
(H₂/CO:1)

(C) Methanol from Combined Feedstocks

Combined Feedstock Concept Compariso
Figure 2-1

The heat required by the gasification reaction is provided by burning lignite within the gasifier. As a result, the concentration of CO_2 is generally quite significant leading to a much lower value of R than the theoretical value of 1. Actual values of R range from about 0.2 to 0.6 and, therefore, there must be even greater shift conversion of CO than would be implied by the theoretical ratio of 1.

2.3.4 Mixing the Two Streams

When mixing the two streams, considerable amounts of steam reformed gas must be used to compensate for the low R ratio of the gasifier's output gas and bring the mixture's ratio to the required $R = 2$.

In some instances, the natural gas requirements are so high that the plant resembles a gas based plant employing little carbon based co-feedstock.

Another approach consists of increasing the gasifier R ratio by eliminating some of the CO_2. Inclusion of a CO_2 removal unit decreases the natural gas requirements and increases proportionately the need for carbon based feedstock. But the overall efficiency of the process decreases since some of the carbon value is vented to the atmosphere.

In the case of relatively cheap lignite, it may prove economically attractive to accept some loss in efficiency by including the CO_2 removal unit, thereby reducing the natural gas consumption.

2.4 ELECTROLYTIC HYDROGEN AS A CO-FEEDSTOCK

In this co-feedstock approach, pure hydrogen, from water electrolysis, is added to the hydrogen deficient raw gas produced through gasification of lignite.

Here, some treatment of the gasifier's output gas is always required to reduce its CO_2 content and to raise its R ratio. If excess CO_2 is present in the synthesis gas, it will regenerate water by reacting with the added hydrogen that was originally produced from water.

$$CO_2 + 3H_2 \longrightarrow CH_3OH + H_2O$$

Figure 2-2 illustrates the electrolytic option. The main technical advantage is that the oxygen by-product generated by electrolysis in the gasifier can be fully used, eliminating the need of an air separation plant to produce oxygen.

2.5 THERMAL EFFICIENCY AND CARBON CONVERSION

Thermal efficiency (TE) and **carbon conversion** (CC) are good indicators of the technical performance of a process.

2.5.1 Thermal Efficiency

Thermal efficiency is the ratio of the energy content of all products to the energy content of all inputs - feedstocks, fuels and power. There is no standard way of expressing the various energy contents. Some sources use the **higher heating value** (HHV), while others use the **lower heating value** (LHV) for feedstocks and fuels. Electric power is converted into its heat content at various conversion efficiencies. Throughout this study all heating values are in (HHV) and electric power is converted at 33% efficiency (1 MWhr = 10.8 GJ).

2.5.2 Carbon Conversion

Carbon conversion is the ratio of the carbon contained in methanol to the total carbon input in the process - feedstocks and fuels. Fluctuations in carbon conversion generally follow corresponding variations of thermal efficiency. If CO_2 is vented to the atmosphere, the corresponding loss of carbon conversion will be reflected in increased consumption of carbon based feedstock and, therefore, there will be a decrease in thermal efficiency.

When electrolytic hydrogen is used, the opposite may occur.

2.6 PROCESS EFFICIENCY AND ECONOMIC VIABILITY

There is an inherent inefficiency in the natural gas to methanol process, due to the rejection of valuable hydrogen and its use as a fuel in the reformer.

The thermal efficiency of this process is increased by adding CO_2 to the synthesis gas. If a CO_2 source is not readily available, for example from a nearby ammonia plant, lignite can be gasified to generate a CO/CO_2 rich gas. However, by doing so, the desired increase in thermal efficiency is more than offset by other energy consuming operations such as feedstock drying and oxygen feed to the gasifier.

Although thermal efficiencies and carbon conversion ratios are lower, the use of lignite, with or without natural gas or electrolytic hydrogen, could be more economically attractive than the gas to methanol process, depending on feedstock prices.

There is no direct relationship between the technical efficiency of a processing option and its economic attractiveness. Therefore, throughout this study, the technical and economic features of each option are evaluated in parallel.

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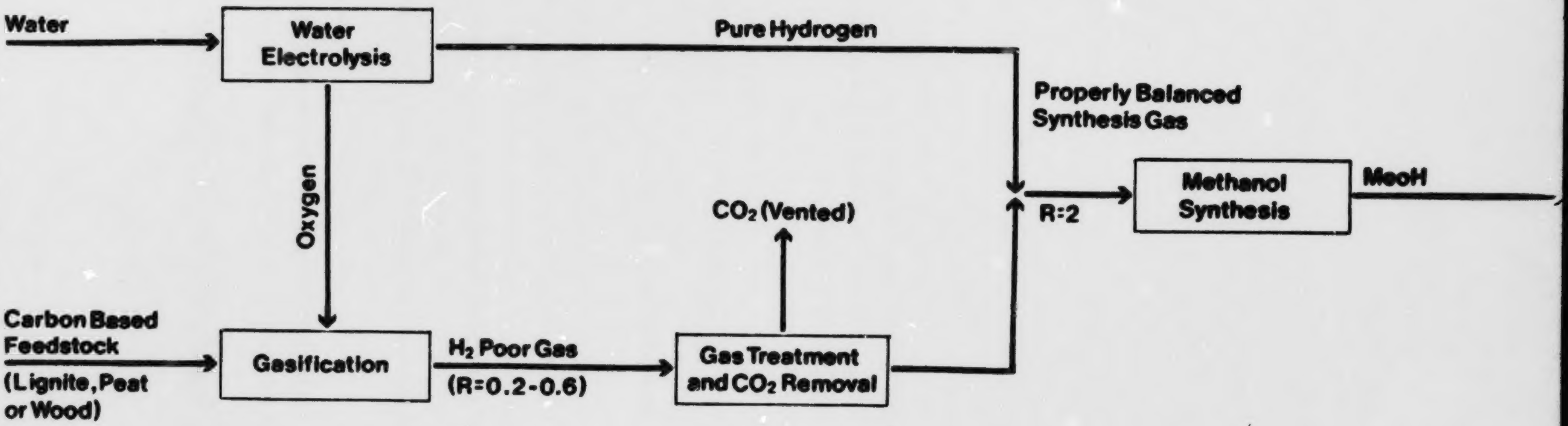

Water
Water Electrolysis
Pure Hydrogen
Oxygen
Properly Balanced Synthesis Gas
Methanol Synthesis
MeoH
R:2
CO₂ (Vented)
Carbon Based Feedstock
(Lignite, Peat or Wood)
Gasification
H₂ Poor Gas
(R=0.2-0.6)
Gas Treatment and CO₂ Removal

The Electrolytic Hydrogen Optic
Figure 2-2

3– METHOD USED FOR ECONOMIC EVALUATION

The basic objective of this analysis is to determine the economic viability and timing of methanol production in Ontario using indigenous lignite resources. Since natural gas is the feedstock used in most existing methanol plants, and could be an additional option for Ontario, we used the economics of natural gas based production as the reference case for assessing viability.

There are a number of gasification technologies that could be considered. In contrast to natural gas technology, commercial performance and economic data is unavailable. It was important, therefore, to screen the technologies in terms of their stage of development and likely technical performance so that more detailed assessments and cost estimates could be restricted to a manageable set of processes.

But, with four gasifier technologies to consider, as well as natural gas and electrolytic hydrogen co-feedstock processing options, the number of alternatives remained large. In addition, the economics of all of the process options could vary substantially depending on feedstock prices, capital cost, market prices for methanol and other factors. We are dealing with plants that would not be built for several years and all of these factors are uncertain. We, therefore, wanted to be able to test the sensitivity of the economics to different assumptions about critical factors. This made our analysis even more complex.

Given the number of alternatives, we used a sequential evaluation process in which the number of options could be reduced and the detail increased at each step. In the initial evaluation stages, the analysis is static. A dynamic discounted cash flow analysis is then introduced for a narrowed range of options to capture the effects of time on both cost and revenue streams.

The other key feature of the analysis is the use of linear equations to express the methanol price that must be realized at the plant gate to cover all capital and operating costs, including a return on investment. In these **realization price** equations, feedstock costs and a **capital recovery factor** (CRF) are treated as parameters that can be varied to:

- o test the sensitivity of the plant gate methanol price to changes in these parameters for a given process;
- o determine the feedstock cost or capital recovery factor that will make one process equivalent to another in terms of the plant gate methanol price.

We used the equation approach in order to properly compare alternatives that use the same or different feedstocks and have significantly different capital costs.

Feedstock cost is one of the most important elements determining viability. With the equations, the effects of feedstock costs, outside the ranges we have assumed, can be readily tested.

The same is true for capital recovery factors. Capital recovery is particularly important when comparing projects that have significantly different capital costs. As a result of the capital recovery consideration, a low capital cost project that has high production costs per unit of output might turn out to be more viable than a high capital cost project with lower production costs per unit of output.

The sequential evaluation process is shown schematically in Figure 3-1.

> **Capital and Operating Costs** were estimated for every option and provided the basis for subsequent steps.
> **Realization Price Linear Equations** were then developed to provide a basis for static comparative evaluations in which feedstock costs and capital recovery factors could be varied.

Discounted Cash Flow Analysis (DCF) adds a time dimension and provides a dynamic view of the project, taking into account projected sales. Project construction schedules, price escalation rates and various plant utilization factors in the early years of operation are included. This provides a set of **internal rates of return** (IRR) and, if required, **net present values** (NPV) for various discount rates.

Preliminary Financial Analysis is performed only on the most promising option. It incorporates typical fiscal and financing elements to illustrate the potential return on equity (ROE) after meeting debt servicing requirements. We emphasize that the financing scheme is not intended to portray the actual financial structure of a specific project. It is only presented as an example to show the economic effects when a project is partly debt funded and taxes are taken into account.

The rest of this chapter explains each of these evaluation steps in detail.

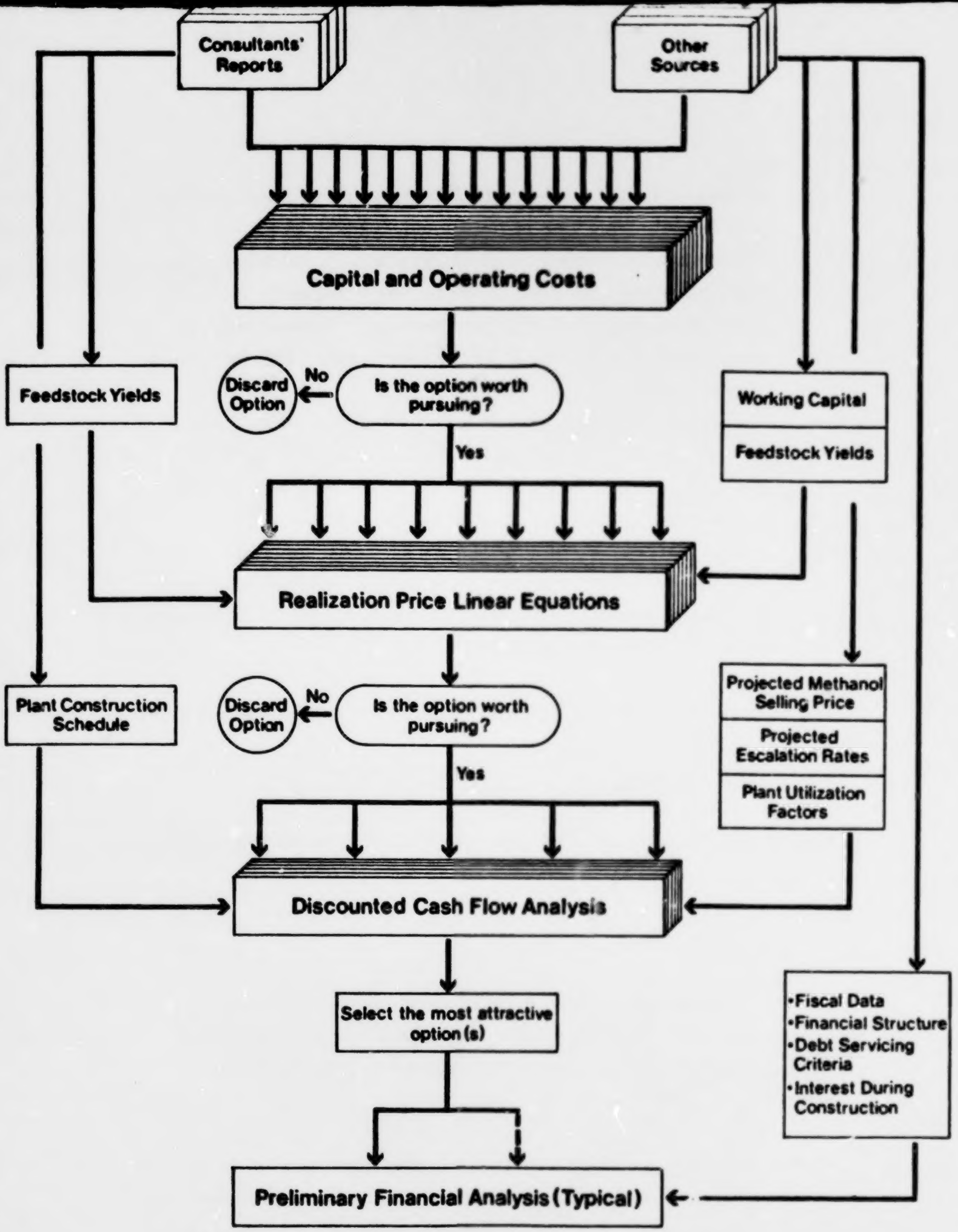
3.1 STEP I: CAPITAL AND OPERATING COSTS

Capital and operating costs are expressed in fourth quarter 1983 Canadian dollars (4Q 1983 Cdn $).

3.1.1 Capital Cost

Alternative processes were investigated mostly at the conceptual engineering level. Block flow diagrams, accompanied by approximate mass and energy balances, are the main support to a **factored cost estimate**, accurate to within 30 to 40%.

When a plant cost is derived from the estimated cost values of its various components, the overall accuracy range remains the same as the average accuracy range for its components. The probability that the **total plant**



Consultants' Reports
Other Sources
Capital and Operating Costs
Feedstock Yields
Discard Option
No
Is the option worth pursuing?
Yes
Working Capital
Feedstock Yields
Realization Price Linear Equations
Plant Construction Schedule
Discard Option
No
Is the option worth pursuing?
Yes
Projected Methanol Selling Price
Projected Escalation Rates
Plant Utilization Factors
Discounted Cash Flow Analysis
Select the most attractive option (s)
•Fiscal Data
•Financial Structure
•Debt Servicing Criteria
•Interest During Construction
Preliminary Financial Analysis (Typical)

Method for Economic Evaluation
Figure 3-1

cost (TPC) will be outside this accuracy range, however, is far smaller than if an estimate were made without considering components separately. Therefore, the breaking down of plant costs into areas, subareas and in some instances, major pieces of equipment, makes the overall cost estimate more accurate.

Capital costs, for each option, are broken down as:

- total installed equipment cost, derived from the various components of the plant
- engineering and construction management costs
- contingencies, including uncertainty factors for technologies which are not yet at the stage of full commercial development
- start-up costs to allow for plant commissioning and initial spare parts inventory.

All of these items constitute the depreciable investment or total plant cost.

3.1.2 Plant Operating Costs

Plant operating costs include the annual averages of all plant expenditures, other than the cost of the feedstocks. The costs of lignite, natural gas and, in the case of the electrolytic hydrogen option, electric power were considered separately.

Operating costs for each option are summarized under these headings.

- total manpower costs, including operating and maintenance labour, fringe benefits, supervision, overhead administration charges

- maintenance parts, catalysts and chemicals
- utilities, other than fuel, such as water, power and ash disposal and

- property tax and insurance

A more detailed breakdown can be found in the KILBORN'S report contained in the appendix. The level of accuracy of the operating cost estimates is considered compatible with that of the capital cost estimates.

3.2 STEP II: REALIZATION PRICE OF METHANOL

From the results obtained in Step I, we eliminated some alternatives for which a better option existed.

Feedstock yields and working capital estimates, obtained from KILBORN'S report and other sources, together with basic cost data established in Step I, provide the basis for the realization price equations.

The realization price of methanol is expressed in cents per litre (4Q, 1983 Cdn $) at the plant gate.

3.2.1 Feedstock Yields (Consumption)

Every processing option is characterized by a set of feedstock consumption values per unit of methanol produced.

As an example, the lignite - natural gas process requires 13.9 **standard cubic feet** (SCF) of gas and 1.19 kg of lignite at 45% moisture per litre of methanol produced.

The feedstock cost component has a substantial impact on the total production cost. Given the wide range of current and projected feedstock prices, we treated prices as equation parameters so that prices could be varied to test sensitivity. Assuming that the price of lignite is L dollars per Tonne ($/T) and the price of gas is G dollars per thousand standard cubic feet ($/MSCF) the feedstock cost component of methanol in the lignite-gas process is:

$$(1.19 \times \frac{L)}{1000} \qquad + \qquad (13.9 \times \frac{G)}{1000}$$

Cost of lignite feedstock (in $/litre of methanol)	Cost of gas feedstock (in $/litre of methanol)

or

$$(0.119L + 1.39G) \text{ in ¢/litre of methanol}$$

In the case where electrolytic hydrogen is the co-feedstock instead of natural gas, the co-feedstock price is expressed in terms of electric power rates E (in ¢/kWhr) rather than hydrogen cost.

The generalized expression for calculating total feedstock and co-feedstock cost components, in cents per litre of methanol (¢/litre), is:

Total feedstock and co-feedstock cost component = $100\,\mathbf{aL} + 100\,\mathbf{bG} + \mathbf{cE}$
where:

a = main feedstock unit consumption in Tonnes of lignite, required per litre of methanol
b = natural gas co-feedstock unit consumption in MSCF/litre of methanol
c = electrolytic hydrogen unit consumption in equivalent kWhr of electric power required per litre of methanol
L = unit price of the main feedstock, lignite, in $/Tonne
G = unit price of natural gas in $/MSCF
E = power rate in ¢/kWhr

All these costs are in constant 1983 Canadian dollars and the factor 100 is included to convert dollar values to cents.

3.2.2 Non-Feedstock Operating Cost

The total annual operating cost, as determined in Step I (in cents per year), is divided by the total annual methanol production (in litres per year) to obtain the operating cost component (in cents per litre).

Non-feedstock operating costs are assumed to be fixed costs, even though some operating cost items are actually variable costs, such as utilities and chemicals. In most cases, however, the variable costs are relatively small in comparison with the total operating cost, so that the approximation is justified.

The total annual production is based on a fully utilized plant, operating at design capacity for an equivalent number of **on-stream days per year** (SDPY). In all cases, the value of SDPY was assumed at 320 days per year for the lignite based options and 330 days per year for the conventional gas based process.

3.2.3 Capital Recovery Factor (CRF)

Capital related charges are intended to cover plant depreciation, interest payments and gross profit. We have followed the common practice of not separating these items at an early stage of project evaluation. At this point, they appear as one number, directly related to the total investment.

Working capital is added to the total plant cost, calculated in Step I, to arrive at the total investment. Working capital is assumed to be sufficient to cover all costs of one and a half months' full plant production.

A capital recovery factor (CRF) is then applied to the total investment to account for these annual capital related charges.

Total Plant Cost + Working Capital = Total investment

Total Investment x CRF ÷ Annual Production = Capital Charges (¢/litre)

Methanol Realization Price (¢/litre) = Feedstock Cost (¢/litre) + Operating Cost (¢/litre) + Capital Charges (¢/litre)

The question often arises as to which CRF value should be applied. Values range widely -- from 5% to 25% -- sometimes even in the same price evaluation.

Two criteria should guide the selection of a CRF:

- There should not be a great difference between the calculated realization price and the expected plant gate selling price of the product. If the product can in fact be sold at exactly the calculated realization price, the project can realize a simple return on investment (ROI) equal, in real terms, to the preselected capital recovery factor (CRF).
- There is a direct relationship between the CRF and the more commonly accepted **discounted cash flow return on investment** (DCFROI), when the actual selling price equals the realizaton price and when cash flows are constant over the life of the asset. In theory, if the plant is constructed in less than one year and remains in operation indefinitely, the DCFROI equals the CRF. In fact, the DCFROI is always lower than the CRF.

For a given plant life of 20 operating years, the difference between the CRF and the DCFROI increases with longer construction periods. Table 3-1 illustrates the relationship between DCFROI and CRF. For simplicity, we assumed that the capital outlays are evenly distributed over the construction years, and that both DCFROI and CRF are expressed in real terms, with no allowance for inflation. It is clear from Table 3-1 that a 5% CRF is unrealistic.

TABLE 3-1

DCFROI, For Various Values of CRF,
Based on a 20 Year Operation

Number of construction years	CRF (%)			
	5	10	15	20
1	0.0	7.8	13.9	19.4
2	0.0	7.3	12.8	17.7
3	0.0	6.9	11.9	16.2
4	0.0	6.5	11.2	15.0

For a 3 to 4 year construction period, a 10% CRF actually generates a 6.5 -6.9% DCFROI in real terms, which should be viewed as a reasonable minimum rate unless the project is heavily subsidized or benefits from advantageous financial conditions.

A 20% CRF, on the other hand, translates into an actual DCFROI ranging from 15 to 16.2% in real terms and may be regarded as excessive for a methanol production facility.

3.2.4 Realization Price Linear Equations

Every alternative process can be represented by this equation

$$M = 100\,aL + 100\,bG + cE + d + \frac{\text{(Total Inv.)}}{\text{(Ann.Prod.)}} \times \text{(CRF)}$$

where:

M is the methanol plant gate realization price in ¢/litre
a, b, c are the unit consumptions of lignite, natural gas co-feedstock and electrolytic power in T/litre, MSCF/litre and kWhr/litre

L is the price of lignite, in \$/T (at 45% moisture)
G is the price of natural gas in \$/MSCF
E is the power rate in ¢/kWhr
d is the total of all the non-feedstock operating costs in ¢/litre
Total Inv. is the total investment, including working capital in dollars
Ann. Prod. is the total annual production of methanol in litres
CRF is the capital recovery factor in percentage points

This equation allows us to:

- analyze the sensitivity of the methanol price to variations in feedstocks costs and capital recovery charges - a static sensitivity;
- estimate the feedstock price conditions which must prevail to make an alternative economically attractive;
- determine the price level of a common feedstock will which provide the same realization price of methanol for competing processes - break-even or cross-over point;
- determine the price levels of different feedstocks which will give the same realization price for competing processes.

3.3 STEP III: DISCOUNTED CASH FLOW ANALYSIS

Processing options, which exhibit economic potential under Step II, undergo a DCF evaluation.

There is a clear difference between the CRF and DCF methods. The former determines what the selling price should be to achieve a return equal to the preselected capital recovery factor. The latter shows what the actual return of the project will be, given a set of projected prices.

In the DCF analysis:

- all costs are expressed in constant dollars -- inflation rates have little influence on DCF returns;
- the total plant cost is distributed among the 2, 3 or 4 construction years, according to ratios which approximate the actual capital outlays;

- working capital is added to the last construction year;
- a common 20 year operating life is assumed for all alternatives;
- all cash flows are based on gross profits before tax;
- cash flows are discounted to the first year of project implementation, thus implicitly accounting for what is commonly known as interest during construction;
- utilization factors lower than 1.0 are applied in the first year(s) of production because plants usually operate below design capacity when they are first commissioned;
- operating costs, other than utilities, are assumed to be independent of plant production levels and are treated as fixed costs.

The DCF analysis includes annual escalation factors applied to:

- wages and salaries
- maintenance parts
- utilities
- tax and insurance
- lignite price
- natural gas plant gate price
- power costs, at plant gate for electrolytic option only
- projected methanol selling price at plant gate

These escalation factors are selected to conform with projections made by the Ontario Ministry of Energy, the National Energy Board, Ontario Hydro and others.

In the discounted cash flow analysis:

- the sensitivity of the project to various price/cost projections, in terms of both price levels and escalation factors, can be tested - dynamic sensitivity;
- the discounted cash flow return on investment (DCFROI) is generally referred to as internal rate of return (IRR) to distinguish it from simple return on investment (ROI);

- the net present value (NPV) of the project can be provided for various discount rates.

3.4 STEP IV: PRELIMINARY FINANCIAL ANALYSIS

In this step, the option which seems potentially most attractive in terms of pre-tax IRR and NPV undergoes a preliminary financial analysis. Our model captures typical fiscal and financing considerations by taking into account:

- the overall financing philosophy for this type of project in Ontario
- probable ranges for debt to equity ratios
- anticipated long-term borrowing rates, the role of commercial bank financing, and the potential for soft loans and/or government subsidies
- the fiscal structure, in terms of depreciation method, tax rate and deferred taxes
- a preliminary forecast of debt payment schedules.

The outcome is an evaluation of possible returns on equity (ROE).

4– METHANOL FROM NATURAL GAS

Because natural gas is used worldwide to produce methanol, the gas-to-methanol (GTM) process, specifically the ICI low pressure process, is used as the reference process in this study. See Figure 4-1.

4.1 PROCESS SUMMARY

Desulphurized natural gas is mixed with steam and reformed at about 870°C, generating a mixture of carbon oxides, primarily CO, and hydrogen.

The heat required in the reforming reaction is provided by purge gas from the methanol synthesis loop, supplemented by natural gas fuel. Synthesis gas ($CO + H_2$) produced in the reformer is compressed to about 80 bars and reacts in the catalytic converter to produce methanol.

The liquid phase containing methanol is separated from the gas and purified by distillation. For fuel grade methanol production, only one distillation column is used to ensure that water content remains below 0.05%.

A portion of the recycled synthesis gas is purged for use as reformer fuel, and the balance is recompressed before re-entering the catalytic converter.

Throughout the process, but particularly in the reformer area, extensive heat recovery systems are used. These include a waste heat boiler and steam superheat; boiler feed water preheat; natural gas preheat; combustion air preheat; reformed gas low level heat used in the distillation reboiler; purge gas reheat. As a result, modern methanol plants are mostly energy self-sufficient.

4.2 RECENT DESIGN IMPROVEMENTS

In the late 1960's, a major breakthrough was achieved in reducing the synthesis pressure from 300 to approximately 100 atmospheres. Since then, the introduction of a number of energy recovery systems has further increased the overall thermal efficiency of the process.

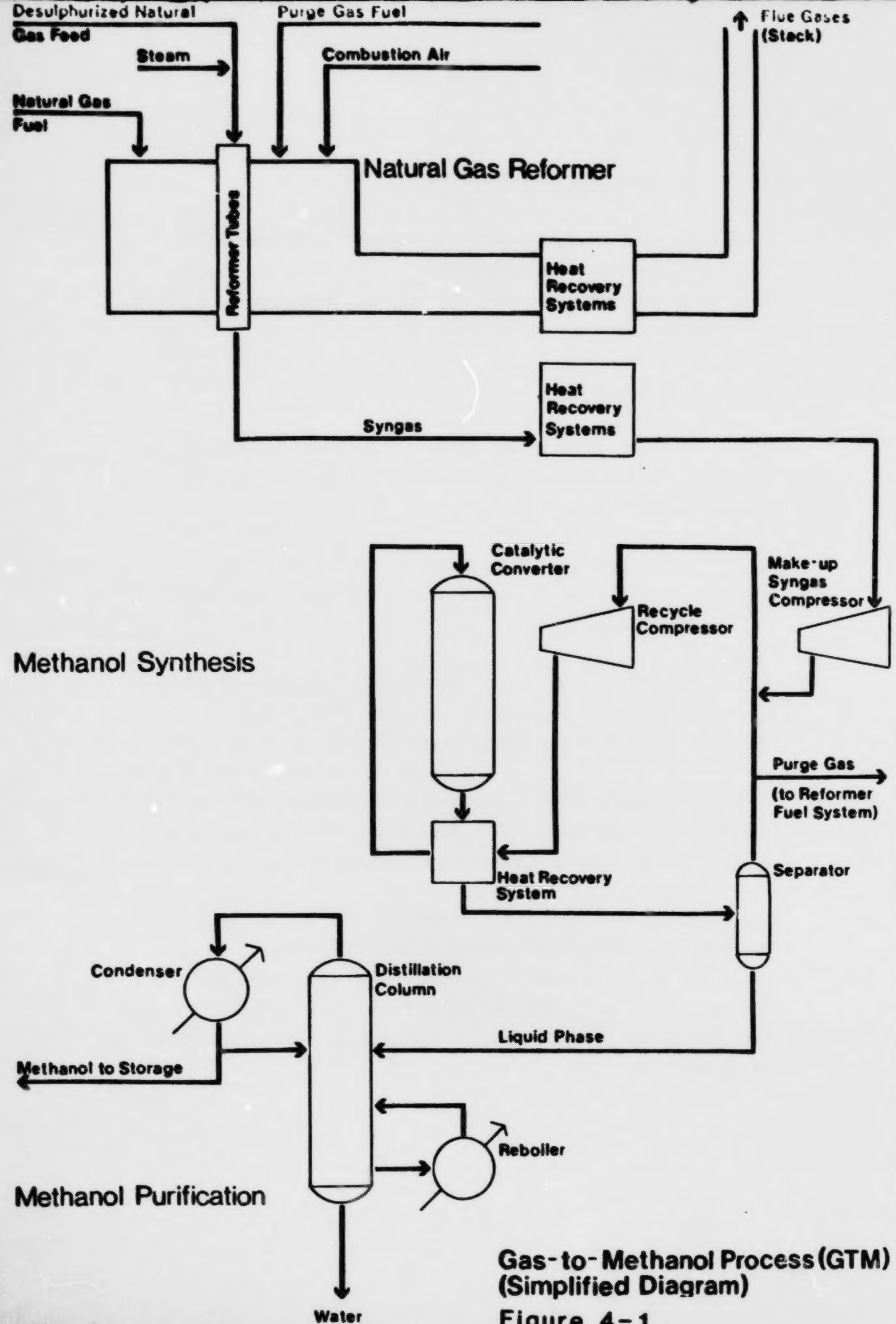
Summary of Improvements

4.2.1 Recovery of waste heat adjacent to the catalytic converter, and increase of the steam generation pressure to 100 bars provided a 9% increase in efficiency.

4.2.2 Improvements in the catalytic converter's quench system, together with a new distillation process design, brought an additional 4% increase in efficiency.

4.2.3 Other factors which produced a further 3% increase in efficiency were: optimization of the synthesis pressure at about 80 bars; introduction of a natural gas saturator, reducing the steam requirement in the reformer; inclusion of a purge gas expander recovering the mechanical energy of this high pressure stream; and, improvements in the design of the reformer. These low energy design improvements are reportedly capable of achieving overall thermal efficiencies in excess of 67%.

Natural gas consumption per unit of methanol produced has consistently decreased over the past few years. Even without considering the lower energy designs, any alternative methanol production option must compete with an efficient gas-to-methanol (GTM) process. As new GTM plants are built, the average process efficiency of plants in operation will continue to improve.


Desulphurized Natural Gas Feed
Purge Gas Fuel
Flue Gases (Stack)
Steam
Combustion Air
Natural Gas Fuel
Reformer Tubes
Natural Gas Reformer
Heat Recovery Systems
Syngas
Heat Recovery Systems
Catalytic Converter
Make-up Syngas Compressor
Recycle Compressor
Methanol Synthesis
Purge Gas (to Reformer Fuel System)
Separator
Heat Recovery System
Condenser
Distillation Column
Liquid Phase
Methanol to Storage
Reboiler
Methanol Purification
Water

Gas-to-Methanol Process (GTM) (Simplified Diagram)

Figure 4-1

4.3 NATURAL GAS CONSUMPTION IN GTM PLANTS

Data provided by the process developer (ICI), engineering firms with methanol expertise, and others is listed in Table 4-1.

Notes: (1) The higher value (27.0 MMBTU/ST) relates to a standard current methanol plant with typical energy recovery systems as shown in 4.2.1 and 4.2.2. The lower value reflects the low energy designs currently available and described in 4.2.3.

(2) The value provided by Chem-Systems refers to a plant starting up in 1983 - 1985 and includes a 6% safety margin over the contractor's design level.

(3) This refers to an existing methanol plant operating in Canada.

(4) The higher value refers to current plant designs (1983), while the lower value can be achieved with "available efficient commercial designs."

A number of units of measurement were used in the documents examined. For consistency, all have been converted into thousands standard cubic feet of gas per Tonne of methanol (MSCF/T) assuming an HHV and a LHV equal to 1000 BTU/SCF and 900 BTU/SCF respectively.

The effect of CO_2 addition at the reformer inlet on the overall gas consumption is shown in some cases. In most methanol plants, however, CO_2 is not available so we do not consider the increased efficiency achieved by this method.

Unless otherwise noted, all gas consumption figures refer to chemical grade methanol (99.85% minimum purity and strict limits for other alcohols and organic compounds). Fuel grade methanol requires less energy than chemical grade methanol because of a decreased requirement in the distillation column reboiler.

TABLE 4-1

NATURAL GAS CONSUMPTION IN GTM PLANTS

SOURCE	REFERENCE DOCUMENT	NAT. GAS CONSUMPTION (as stated)	NAT. GAS CONSUMPTION (MSCF/T)	BASIS	NOTES
IMPERIAL CHEMICAL INDUSTRIES (ICI)	REF (1)	27.0 MMBTU/ST (LHV)	33.08	No CO_2 - Chem grade	(1)
		26.0 MMBTU/ST (LHV)	31.85	No CO_2 - Chem grade	
		25.3 MMBTU/ST (LHV)	30.99	With CO_2 - Chem grade	
HUMPRHEYS AND GLASGOW	REF (2)	7.19 Gcal/T (LHV)	31.70	No CO_2 - Chem grade (based on 2500 TPD plant)	
KELLOGG-RUST SYNFUELS INC.	REF (3)	27.0 MMBTU/ST (LHV)	33.08	No CO_2 - Chem grade	
		26.6 MMBTU/ST	32.59	No CO_2 - Fuel grade	
		26.1 MMBTU/ST	31.97	With CO_2 - Chem grade	
		25.4 MMBTU/ST	31.12	With CO_2 - Fuel grade	
UHDE GmbH	REF (4)	30.56 GJ/T (LHV)	32.19	No CO_2 - Chem grade	
		29.31 GJ/T (LHV)	30.88	With CO_2 - Chem grade	
CHEM-SYSTEMS	REF (5)	8.311 GCal/T (LHV)	32.98	With CO_2 - Chem grade	(2)
CONFIDENTIAL SOURCE		900 Nm^3/T	33.59	No CO_2 - Chem grade	(3)
WORLD BANK	REF (6)	13500 BTU/lb (LHV)	33.08	No CO_2 - Chem grade	(4)
		13088 BTU/lb (LHV)	32.07	No CO_2 - Chem grade	

A close examination of Table 4-1 leads to the following conclusions.

- CO_2 addition reduces the overall natural gas consumption by an average of 3.5% for chemical grade methanol.
- Without considering the very efficient low energy designs, and on the basis of chemical grade with no CO_2 addition, the total gas consumptions are:

ICI	33.08	MSCF/T
KELLOGG	33.08	MSCF/T
UHDE	32.19	MSCF/T
CHEM-SYSTEMS (32.98 x 1.035)	34.13*	MSCF/T
CONFIDENTIAL	33.59	MSCF/T
WORLD BANK	33.08	MSCF/T
AVERAGE	33.19	MSCF/T

* includes a 6% safety margin

- The production of fuel grade methanol requires 1.5% less gas than the chemical grade (Kellogg).

In our study, however, we have based feedstock consumption for the GTM process on chemical grade methanol requirements.

33.19 MSCF/Tonne
or
26.3 SCF/litre (specific gravity of methanol: 0.7928)

This value corresponds to a 64.8% thermal efficiency (HHV). The energy corresponding to the electric power used is negligible.

Alternative processes based on lignite were evaluated on the basis of fuel grade methanol. Their comparison, to the chemical grade GTM process, should then be regarded as conservative.

4.4 CAPITAL AND OPERATING COSTS

An in-depth study of gas-to-methanol plants was not undertaken. A number of these plants are in operation and reliable reference data exists for capital and operating costs.

A standard 1100 TPD methanol plant, including the required offsite facilities, was selected for our cost evaluation.

4.4.1 Capital Cost

Total plant cost (TPC) is derived from basic cost data presented by various sources, as listed in Table 4-2.

We use a TPC of $160 million for 1100 TPD GTM plant. Independently, Monenco completed a preliminary cost estimate and arrived at the same result.

4.4.2 Operating Cost (Excluding Feedstocks)

Based on reports from the World Bank, Davy McKee and Chem-Systems, as well as Monenco's preliminary cost evaluation, we estimate the operating costs of a 1100 TPD methanol plant at $13.1 million/year (4Q 1983 Cdn $), distributed as:

		MM$/Yr
o	Total Manpower	3.6
o	Maintenance Parts, Catalysts and Chemicals	5.4
o	Utilities, other than gas	0.5
o	Tax and Insurance	3.6
o	Total	$ 13.1

TABLE 4-2

CAPITAL COST OF 1100 TPD GAS BASED METHANOL PLANTS

SOURCE	REFERENCE DOCUMENT	PLANT CAPACITY (TPD)	TIME & LOCATION	CAPITAL COST (AS STATED) MMUS$	TOTAL PLANT COST PRORATED MMCdn $ (4Q 1983)*
WORLD BANK	REF (6)	907	2nd Q, 82 Industrial Country	98.5	157.2
DAVY MCKEE	REF (7)	1088	2nd Q, 82 US Gulf	105	163.8
CHEM-SYSTEMS	REF (5)	1200	mid 1980 Alberta	112.4	158.6

*Note: The proration to 4Q 1983 Canadian Dollars for an 1100 TPD plant built in Ontario was based on:

- Capacity/cost relationship based on exponent 0.65
- 1 US$ = 1.23 Cdn $
- 8% average annual inflation rate
- Ontario/US Gulf construction cost ratio : 1.10
- Alberta/Ontario construction cost ratio : 1.10
- 2% start-up costs included in TPC

4.5 REALIZATION PRICE OF METHANOL: GTM PROCESS

4.5.1 Annual Methanol Production

A GTM plant is expected to operate at full capacity for an equivalent of 330 days per year. An 1100 TPD plant will then produce 363,000 TPY of methanol, or 458 million litres annually.

4.5.2 Feedstock Cost

Using the equation from section 3.2.1

$$\text{Total feedstock and co-feedstock cost component} = 100\,aL + 100\,bG + cE$$

and the calculations in 4.3, we arrive at a gas cost of 2.63G cents per litre of methanol (where G is the price of gas in $/MSCF).

4.5.3 Operating Cost

Annual operating costs of $13.1 MM/yr, distributed over 458 million litres, represent 2.9¢/litre.

4.5.4 Total Investment

For a typical gas price of $4.5/MSCF, the total production cost would be:

$$(2.63 \times 4.5) + 2.9 = 14.7\text{¢/litre or } \$204{,}000\text{/day}$$

Working capital, based on one and a half months at full production would then amount to about $9MM. The total investment for an 1100 TPD GTM plant is estimated to be $169 million.

4.5.5 Realization Price Linear Equation

Based on: $M = 100\,aL + 100\,bG + cE + d + \frac{(\text{Total Inv.})}{(\text{Ann.Prod.})} \times (\text{CRF})$

the realization price of methanol, for any given value of gas price (G) and capital recovery factor (CRF) is:

$$M = 2.63G + 2.9 + \frac{169}{458} \times (CRF)$$

or

$$\boxed{M\ (¢/l) = 2.63G + 2.9 + 0.37\ (CRF)}$$

4.6 PRORATION TO LARGE SCALE METHANOL PLANTS: GTM* PROCESS

The cost data developed earlier refers to a standard size 1100 TPD methanol plant, GTM.

Economies of scale can be obtained by increasing plant capacity. Currently, a 2500 TPD size is the largest single train methanol plant that can be designed.

Using information presented in references (5) and (6), cost data of a 2500 plant, built in southern Ontario and operating at full capacity 330 days per year is estimated to be:

Total Plant Cost (including start-up costs)	$ 287 MM
Working Capital	$ 20 MM
Total Investment	$ 307 MM
Total Non-Feedstock Operating Costs	$21.1 MM/yr*
Natural Gas Consumption	26.3 SCF/litre
Annual Production	825,000 TPY (1.04 billion litres/year)

Realization price equation

$$\boxed{M\ (¢/l) = 2.63G + 2.0 + 0.30\ (CRF)}$$

*Breakeven of operating costs:

	$MM/Yr
o Manpower :	4.9
o Maintenance :	10.5
o Utilities :	1.2
o Tax & Insurance :	4.5
o Total	21.1

5— METHANOL FROM LIGNITE

5.1 LIGNITE RESOURCES IN ONTARIO

Ontario's lignite deposits are located in the James Bay lowlands. The lignite resources have been surveyed since the 1920's and exploration programs are still underway over the western James Bay area. The focus of attention, until recently, has been the Onakawana lignite deposits, with proven reserves of 180 million Tonnes.

But other deposits may offer even greater potential. For example, Onexco Minerals Ltd. have completed three years of exploration. Based on their results, in addition to the Onakawana deposits, geological consultants estimate a regional potential for the lowlands of 700 million Tonnes. Their predictions are encouraging, but are not yet proven. Therefore, no final assessment can be made with regard to mining possibilities and lignite quality.

5.1.1 Previous Studies
Onakawana Lignite Deposits

The Onakawana lignite deposits are by far the best known and our study deals only with the possible use of these deposits for the production of methanol. As summarized in Table 5-1, there have been a number of reports over the years dealing with the potential uses of Onakawana lignite. They include geological surveys, burn tests, and the potential for gasification, drying and briquetting, and power generation.

The proven reserves of 180 million Tonnes can easily sustain more than one large scale methanol plant for a 20 year operation.

Saskatchewan Lignite Deposits

Considerable similarities exist between Saskatchewan and Onakawana lignites. Some information, supplied by Saskoil, on Saskatchewan lignite was used in our study.

5.1.2 Typical Lignite Analysis

Obtaining absolutely representative samples of lignite over a wide geographic area is difficult. However, we believe that the samples analyzed are sufficiently representative for the purposes of this study.

Two samples of Onakawana lignite were analyzed for an Ontario Ministry of Energy study of direct liquefaction of Onakawana lignite for the production of liquid hydrocarbons. For consistency, we used the same analyses.

The Canada Centre for Mineral and Energy Technology (CANMET), in Ottawa, determined proximate and ultimate analyses, ash composition and calorific value. Petrographic analyses were done by the University of Regina.

Table 5-2 summarizes the most significant average results. These were used as the basis for all material and heat balances.

The Ontario Energy Corporation, under a joint program co-sponsored by CANMET, subcontracted to Kilborn Ltd. major portions of the lignite to methanol study. Kilborn's final report is in Appendix I.

TABLE 5-1

PREVIOUS WORK ON ONAKAWANA LIGNITE

REPORT DESCRIPTION	PREPARED BY
Report on Lignite Burn Tests, July 1980	Shawinigan-Steag Comp. (SSC)
Pilot-Scale Combustion Trials with Onakawana Lignite (1980-81)	CANMET
Combustion Performance on Onakawana Lignite in Pulverized - Fired Research Boiler - General Description, September 1967	CMRC
Pilot Scale Combustion - Evaluation of Manalta Briquettes, October 1978	CANMET
Memorandum: Suitability of Coal, Samples as a Fuel for Koppers-Totzek Gasification, September 1980	KRUPP-KOPPERS
Methanol From Onakawana Lignite, September 1980	Edward Werner Cook, Inc. Inc.
Lignite Conversion Study Task III - Lignite Gasification Processes, August 1980	Partec-Lavalin/ Ebastic Lavalin, Inc.
Comparison of Selected Gasification Processes, January 1981	Ebastic Lavalin, Inc.
Preliminary Study Commercial Development of Methanol From Onakawana Lignite, February 1981	Peter Turner & Associates
Heavy Water Complex Feasibility Study, May 1969	H.G. Acres & Co. Ltd.
File: Lignite Quality 1930 - 1972	Babcock/Department of Mines
The Estevan Dryer and Onakawana Lignite Briquetting Study, October 1978	MCL
Order-of-Magnitude Cost Estimates of Drying and Briquetting Plants for Onakawana Lignite, December 1974	Techman
Onakawana Development Study Volume II - Mine Development	SSC
Review of Drilling Programs ODL	Wierco
Review and Proposals on Production of Liquid Fuels From Onakawana Lignite, May 1981	Tapics
Mine Development Study for Onakawana Phase II	Onakawana Development Limited

TABLE 5-2

ONAKAWANA LIGNITE ANALYSIS SUMMARY

		%
Average Estimated Moisture Content	as received basis	45.00
Carbon	dry basis	59.30
Hydrogen		3.81
Sulphur		1.41
Nitrogen		0.71
Ash		18.81
Oxygen (by difference)		15.96
TOTAL		100.0%

Heating Value (HHV)
(dry basis) 5337 Kcal/kg
or
22.33 GJ/T

5.2 LIGNITE GASIFICATION TECHNOLOGIES

5.2.1 Previous Investigations on Onakawana Lignite

In the 1970's several gasification technologies were evaluated for Onakawana lignite, particularly the Koppers-Totzek, Winkler and Texaco gasifiers.

Onakawana lignite was found to be suitable for the production of either medium BTU gas or methanol. The atmospheric Winkler gasifier was singled out as the most economically attractive technology[1].

5.2.2 Pre-Selection of Potential Gasifiers

By December 1982, a comprehensive study carried out by Saskoil on the gasification behaviour of the Saskatchewan lignite was almost complete. In addition to the evaluation of established gasification technologies, such as Lurgi and Koppers-Totzek, new second generation high temperature, pressurized systems were also considered. These were the High Temperature Winkler (HTW) and the Westinghouse gasifiers.

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[1] Comparison of Selected Gasification Processes, January 1981
Ebastic Lavalin, Inc

The pre-selection of potential gasifiers, likely to be suitable for Onakawana lignite, was based largely on the Saskoil study of Saskatchewan lignite. A fifth gasifier, Texaco, was initially included in the evaluation. A preliminary examination, however, eliminated the Texaco gasifier because of the difficulties in slurrying the high moisture Onakawana lignite.

5.2.3 Preliminary Results

Phase Zero (0) of the lignite-to-methanol study carried out by Kilborn, established these conclusions.[2]

- All four gasifiers, Lurgi, Koppers-Totzek, HTW, Westinghouse, were found technically acceptable to process Onakawana lignite into a synthesis gas suitable for methanol production.
- However, there were two drawbacks to the Lurgi gasifier. It was not able to process lignite fines and it produced a number of undesirable by-products because it operated at relatively low temperatures.

5.3 SELECTION OF THE MOST PROMISING GASIFIER[3]

5.3.1 Basis for Evaluation

Typical characteristics of Onakawana lignite were supplied to all the gasifiers' developers, who responded with their best preliminary assessment of the gasification parameters. They provided basic data on:

[2] Kilborn Study, Phase 0, Appendix I

[3] Kilborn Study, Phase I, Appendix I

- raw gas composition
- oxygen and steam consumption
- operating pressure and temperature
- ash disposal and lignite feed systems
- preliminary capital cost

5.3.2 Processing Options

Several options were evaluated.

- Lignite alone, considered as the base case
- Lignite + natural gas reforming
- Lignite + natural gas reforming and hydrogen separation through the Pressure Swing Absorption System (PSA)
- Lignite + electrolytic hydrogen
- Power co-generation vs purchased power

5.3.3 Results

These are the results of the preliminary evaluation.

- The separation of the hydrogen in the reformed gas stream through the PSA system did not show any advantage over the standard natural gas co-processing option.
- Power co-generation in the electrolytic hydrogen option did not appear as a viable option, considering the expected purchased power rates.
- The considerable amount of fines generated by the Lurgi process, made this process impractical since no use can be made of this material. The Lurgi process was, therefore, eliminated.
- The cancellation of the Westinghouse demonstration plant in South Africa made its commercialization unlikely before 1990.

However, the Westinghouse gasification technology is likely to be acquired by Kellogg-Rust Synfuels Inc. New developments may be forthcoming.

- The Koppers-Totzek technology is well established and could be used in the gasification of Onakawana lignite.
- The High Temperature Winkler (HTW) had a slightly better economic performance than Koppers-Totzek and showed potential for commercial use within a few years.

Therefore, the HTW gasifier was selected for further investigation.

5.4 EVALUATION OF THE HTW GASIFIER

5.4.1 Gasification Tests

Bench-scale tests were performed at the Rheinbraun's process development unit in Aachen, West Germany. Both Onakawana and Saskatchewan lignite samples were tested, the latter to provide some indication of the impact of variations in chemical composition on the gasification behaviour.

The full test report, excluding confidential material, is in Appendix 2. Preliminary indications derived from these short duration tests, performed on a small lignite sample (300 kg), indicate that Onakawana lignite is well suited for fluidized bed gasification in spite of its relatively high ash content and low ash fusion temperature.

These tests provided more reliable values for yields, steam and oxygen consumption, and raw gas composition.

These results were projected to commercial scale application by Uhde GmbH Dortmund, West Germany who hold the exclusive licence for the HTW technology. Their projections were also based on extensive pilot testing of German lignite (Rheinische Braun Coal).

Table 5-3 summarizes the projected HTW gasification conditions of Onakawana lignite.

TABLE 5-3

HTW GASIFICATION OF ONAKAWANA LIGNITE

Oxygen consumption (Nm^3/kg lignite maf)		0.51
Steam consumption (kg/kg lignite maf)		0.30
CO_2 recycle (Nm^3/kg lignite maf)		0.125
Lignite feed moisture content (%)		12
Carbon conversion (%)		95
Specific synthesis gas yield (Nm^3 ($CO + H_2$)/T lignite maf)		1655
Raw gas composition (% vol) (H_20 and N_2 fee)		
H_2	33.0	
CO	44.5	
CO_2	19.9	
CH_4	2.0	
H_2S	0.6	
Total	100.0%	

5.4.2 Overall Lignite-To-Methanol Processes

Based on the encouraging, preliminary results of the HTW gasifier, we established an overall lignite-to-methanol design[4] assuming that:

- the plant is located in the Onakawana area, at the mouth of the mine, to avoid costly transportation of a high moisture feedstock;[5]
- the design is based on the largest, **single train**, low pressure ICI methanol plant of 2500 Tonnes per day. A plant of this size takes full advantage of economies of scale. Some engineering firms have reportedly studied the feasibility of a single train 7000 TPD methanol plant. But none of these plants have been built.

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[4] Kilborn Study, Phase 2, Appendix I

[5] Because methanol has to be transported to southern Ontario markets, there may be some economic advantages in shipping pre-dried lignite by rail to a plant built in Southern Ontario.

An overall energy balance was completed at the conceptual engineering level. This assumed that all the power requirements of the major compressors are provided by high pressure steam raised in the waste heat boilers. This is supplemented if necessary by a lignite-fired auxiliary boiler. Medium and low pressure steam used in the process is generated in back pressure turbines. It is worth noting that, given the relative price range of lignite and purchased electric power, there is no real advantage in using large electric motors rather than condensing steam turbines. This explains why, at this stage of the study, a few condensing turbines are shown in the heat balance diagrams.

Three cases were examined in terms of feedstock combinations.

- The lignite-to-methanol case (LTM) referred to as the base case, uses only lignite as a feedstock.
- The lignite and natural gas option (LGTM) is based on steam methane reforming without hydrogen separation through the PSA system. The relative price range of natural gas and lignite is such that raw gas downstream of the gasifier is enriched in hydrogen (CO shift reaction) and reduced in its CO_2 content, before mixing with the reformed gas. This approach is explained in section 2.3.4.
- The lignite and electrolytic hydrogen option (LHTM) is based on partial hydrogen addition. The water electrolysis unit is designed on the basis of the oxygen required by the gasifier. The final hydrogen concentration of the synthesis gas is achieved through partial CO shift conversion and CO_2 removal, followed by electrolytic hydrogen addition.

5.5 CAPITAL AND OPERATING COST ESTIMATES

5.5.1 Capital Costs

Projected capital costs of the three processing options are summarized in Table 5-4. All costs are in million 4Q 1983 Cdn $.

TABLE 5-4

CAPITAL COSTS OF A LIGNITE BASED METHANOL PLANT (2500 TPD)

	LTM	LGTM	LHTM	NOTES
Total Installed Equipment Cost	659.6	445.9	606.1	(1)
Engineering & Construction Management	117.6	79.5	106.1	
Contingencies	82.3	55.7	74.3	
Start-up Cost	17.2	11.6	15.7	(2)
Total Plant Cost (TPC)	876.7	592.7	802.2	

NOTES (1) A high voltage power line is included in the cost of the LHTM option. The cost of the required natural gas pipeline is not included in the LGTM option. This item is included in the operating costs and incorporated in the methanol product cost.

(2) Start-up costs are estimated at 2% of the plant cost.

5.5.2 Operating Costs

Non-feedstock operating costs are summarized in Table 5-5. All costs are in million 4Q 1983 Cdn $ per year.

TABLE 5-5

OPERATING COSTS OF A LIGNITE BASED METHANOL PLANT (2500 TPD)

	LTM	LGTM	LHTM	NOTES
Total Manpower Costs	11.6	11.6	11.6	(1)
Maintenance, Catalysts & Chemicals	16.5	11.0	10.9	
Power	3.6	3.0	-	(2)
Property Tax & Insurance	5.9	4.1	5.3	
Total Operating Costs	37.6	29.7	27.8	

NOTES (1) The total manpower costs are derived from Kilborn's cost estimates and further discussions with Manalta Coal. The average salary per hourly paid employee (114 operating + 90 maintenance people) is about $46,300/year, including fringe benefits. Added to this cost are all overhead and administration expenses, amounting to over $2 million per year.

(2) Power costs for the LHTM option are included in the feedstock costs.

5.6 REALIZATION PRICE OF METHANOL

5.6.1 Annual Methanol Production

The three lignite-to-methanol options LTM, LGTM and LHTM, are based on a capacity of 2500 TPD.

We assumed, for the three options, a constant on-stream rate of 320 days per year at full capacity. The total annual methanol production is then 800,000 Tonnes per year, or slightly over one billion litres.

5.6.2 Feedstock Consumption

Material and heat balances lead to the projected consumption of lignite, natural gas and electrolytic hydrogen. The latter co-feedstock is expressed in terms of equivalent electric power requirements. Table 5-6 summarizes the feedstock consumption per Tonne of methanol for the three options, and indicates the thermal efficiency and carbon conversion of each.

TABLE 5-6

FEEDSTOCK CONSUMPTION IN LIGNITE BASED PLANTS

	LTM	LGTM	LHTM
Plant Production (TPD)	2500	2500	2500
Lignite Consumption TPD moisture free (mf)			
Feed	4261	1447	2636
Fuel	1202	616	657
Total	5463	2063	3293
Natural Gas Consumption (TPD)			
Feed	-	727	-
Fuel	-	143	-
Total	-	870	-
Electric Power (mWhr/day)	-	-	10128
Overall Unit Consumption			
Lignite T/litre (mf)	0.001732	0.000654	0.001044
Gas MSCF/litre	-	0.01386	-
Power kWhr/litre	-	-	3.211
Thermal Efficiency (%)	46.7	61.2	31.3
Carbon Conversion (%)	28.9	50.6	48.0

5.6.3 Feedstock Cost

The price of lignite is based on an as received basis. We assume here, for simplicity, a constant moisture content of 45%.

Using Table 5-6, with the equation developed in 3.2.1.

Total feedstock and co-feedstock cost component = 100 aL + 100 bG + cE

we arrive at the feedstock cost component of lignite based methanol in cents/litre.

LTM	0.315L
LGTM	0.119L+ 1.39G
LHTM	0.189L + 3.21E

where L = price of lignite, at the plant gate (in $/Tonne, at 45% moisture)

G =price of natural gas, excluding any fixed charges related to the construction of a new pipeline and its maintenance. (in $/MSCF)

E =power rate, at plant gate (in ¢/kWhr)

5.6.4 Operating Costs

Annual operating costs listed in Table 5-5 distributed over 1.009 billion litres are converted into cents/litre of methanol.

LTM	3.7¢/litre
LGTM	2.9 + 1.1* = 4.0¢/litre
LHTM	2.75¢/litre

* This figure accounts for the pipeline demand charge, as prorated from information obtained from Northern and Central Gas ($10.8 millions/year)

5.6.5 Total Investment

For typical lignite, gas and power prices of $16/T, $5/MSCF and 2¢/kWhr, respectively, the unit production cost would be:

LTM (0.315 x 16) + 3.7 = 8.7¢/litre

LGTM $(0.119 \times 16) + (1.39 \times 5) + 4.0 = 12.8$¢/litre
LHTM $(0.189 \times 16) + (3.21 \times 2) + 2.75 = 12.2$¢/litre

The required working capital, assumed to cover one and a half months of production, is then estimated at $12 MM for LTM and $18 MM for both LGTM and LHTM. These amounts, added to the total plant cost shown in Table 5-4 represent the total investment (in millions of dollars)

LTM	$888.7 MM
LGTM	$610.7 MM
LHTM	$820.2 MM

5.6.6 Realization Price Linear Equations

Using the equation from 3.2.4

$$M = 100aL + 100bG + cE + d + \frac{(\text{Total Inv.})}{(\text{Ann. Prod.})} \times (\text{CRF})$$

and incorporating cost elements already discussed, the final price equations are:

$$\boxed{\begin{array}{ll} \text{LTM} & M(\text{¢}/l) = 0.315L + 3.7 + 0.88 \times (\text{CRF}) \\ \text{LGTM} & M(\text{¢}/l) = 0.119L + 1.39G + 4.0 + 0.60 \times (\text{CRF}) \\ \text{LHTM} & M(\text{¢}/l) = 0.189L + 3.21E + 2.8 + 0.81 \times (\text{CRF}) \end{array}}$$

These equations are discussed in Comparative Economics in Chapter 7.

6— FEEDSTOCK COST PROJECTIONS

This chapter summarizes the cost projections for lignite, electric power and natural gas that are used in comparative economic evaluations of different processing operations.

The projections presented here establish the base for economic comparisons, but sensitivity analyses show the effects of different feedstock cost assumptions.

6.1 LIGNITE

Manalta coal provided estimates of the cost of lignite, based on their mining experience with other coals. These costs are for lignite as received at the methanol plant, with an estimated moisture content of 45% by weight.

Lignite costs will depend on the mining scale.

- For a large-scale mine producing about 3.2 million Tonnes per year, the lignite cost is estimated at $15.75 per Tonne, ±15%. This scale corresponds to the requirements of the lignite to methanol processing option without a co-feedstock (LTM).
- For a smaller scale mine producing about 1.9 million Tonnes per year, the cost is estimated at $19.39 per Tonne, with an uncertainty range of +20% to -15%. The corresponding lignite cost range is $16.48 to $23.27 per Tonne. This scale of mine would meet the requirements of the lignite and electrolytic hydrogen co-feedstock processing option (LHTM).

The large-scale mining cost estimates are reportedly more reliable. Costs for the smaller operation could actually be lower than estimated, but that possibility cannot be confirmed without more detailed studies.
For our analysis, lignite costs are assumed to escalate at 0.5% per year in real terms.

6.2 ELECTRIC POWER

The electrolytic option for lignite, uses large quantities of power. A 2500 TPD methanol plant based on lignite and electrolytic hydrogen would use 420 megawatts (MW).

Conventional power rates do not apply to high levels of industrial power consumption. For example, under Ontario Hydro's blended power class, the user pays 2.0¢/kWhr, without any demand charge. But, this rate is only in effect for eight months a year, and the average annual power rate is actually about 2.2¢/kWhr.

Although this rate won't necessarily apply to the electrolytic options, it does show that a power rate can be negotiated at or below 2.2¢/kWhr. Far lower power rates are being discussed for similar projects in other provinces.

After preliminary discussions with Ontario Hydro, we estimated a base price of 2¢/kWhr. While we do not attempt to define a power rate structure, the implications of different power rates are discussed.

Projections made by Ontario Hydro show an escalation of 0.5% in real terms, to the year 2000.

6.5 NATURAL GAS

In our analysis, the projected price of natural gas is the main factor that determines the time frame for using feedstocks indigenous to Ontario.

Under current Canadian pricing policies, natural gas prices are linked directly to crude oil prices. In contrast to the other feedstocks, projected prices of natural gas show significant increases over the next 20 years.

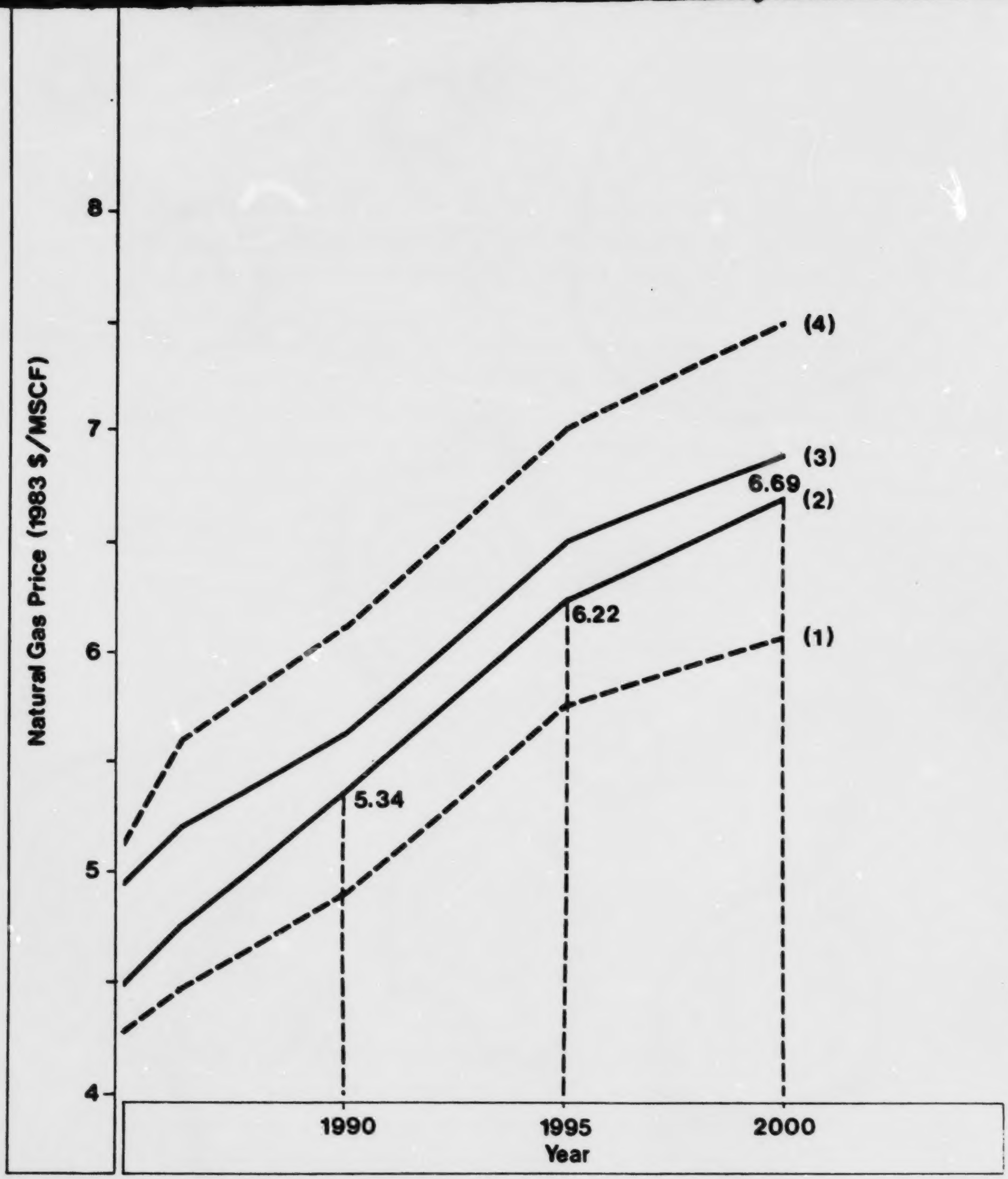
Projections of industrial natural gas rates prepared by the Ontario Ministry of Energy and the National Energy Board are shown in Table 6-1 and Figure 6-1. Both sources predict that, by 1990, Toronto city gate natural gas prices could be close to $5.50/MSCF, in constant 1983 Cdn $.

TABLE 6-1

PROJECTED NATURAL GAS PRICES
(in 1983 Cdn $/MSCF, Toronto)

Year	Ontario Ministry of Energy		National Energy Board	
	Low Case	High Case	Low Case	High Case
1983	4.55	4.55	4.55	4.55
1984	4.32	4.32	4.32	4.32
1985	4.28	4.66	4.72	5.12
1986	4.44	5.04	4.82	5.57
1990	4.91	5.77	5.14	6.10
1995	5.77	6.67	5.99	6.97
2000	6.08	7.29	6.22	7.50

The methanol market analysis, developed in a separate report, indicates that a new methanol production facility won't be required in Ontario before 1990, if then. Therefore, the static comparative evaluation of the various methanol production routes uses natural gas price projections from 1990 on. To be conservative, we used the average of the Ministry of Energy high and low projections.



(1) Low Case, Ontario Ministry of Energy
(2) Average, Ontario Ministry of Energy
(3) Average, National Energy Board
(4) High Case, National Energy Board

Natural Gas Price Projection
Figure 6-1

7 — COMPARATIVE ECONOMICS

The realization price equations are used in this chapter to compare the lignite options to each other and to the large scale natural gas process. Nomographs are presented at the end of the chapter which allow sensitivity testing of various feedstock prices and capital recovery factors for the preferred lignite option as well as for the natural gas option.

The lignite-to-methanol (LTM) option, without a co-feedstock, was found to be the most promising and was carried forward to the discounted cash flow analysis in Chapter 8.

7.1 THE LTM VERSUS THE LGTM OPTIONS

Methanol can be produced in a 2500 TPD plant located in Onakawana, by the LTM and LGTM process options. Plant gate prices are defined by the equations presented in 5.6.6.

$$\text{LTM} \quad M(¢/l) = 0.315L + 0.88\,(CRF) + 3.7$$

$$\text{LGTM} \quad M(¢/l) = 0.119L + 1.39G + 0.60\,(CRF) + 4.0$$

For simplicity, we shall assume that the same lignite price applies to both options.

For a given capital recovery factor (CRF), there is a relationship between the gas cost G and the lignite cost L which makes one option as attractive as the other.

The breakeven cost equation is:

$$0.196L - 1.39G + 0.28\,(CRF) - 0.30 = 0$$

or

$$\boxed{L = 7.09G - 1.43\,(CRF) + 1.53}$$

This equation is represented graphically in Figure 7-1 for capital recovery factors of 10% and 15%. Each straight line shows the breakeven conditions of lignite and gas costs which make the total cost of the two processes, LTM and LGTM, the same. For example, if G = \$5/MSCF and L = \$22.7/T, methanol can be produced by either process for the same realization price, with a common capital recovery factor of 10%.

Each of the straight lines shown in Figure 7-1 separates two areas. The upper left area corresponds to high lignite prices and low gas prices. The lower right area corresponds to low lignite prices and high gas prices.

Any combination of gas and lignite prices which falls into the upper left area would indicate that the LGTM option is more attractive than the LTM option.

Since lignite prices are projected to be about \$16/T and gas price projections average \$5.5/MSCF by 1990, we conclude that the LTM option is more attractive than the LGTM option. This is particularly true in the long run, regardless of the adopted capital recovery factor. Also, lignite prices for the LGTM option are likely to be higher than for the LTM option because the mining scale would be different.

Because the LGTM option offers no economic advantage over the LTM option, it is eliminated.

7.2 THE LTM VERSUS THE LHTM OPTIONS

Here we use the same method as for LTM/LGTM.

$$\text{LTM} \quad M(\text{¢}/l) = 0.315L + 0.88\,(\text{CRF}) + 3.7$$

$$\text{LHTM} \quad M(\text{¢}/l) = 0.189L + 3.21E + 0.81\,(\text{CRF}) + 2.8$$

Assuming the same lignite prices for both options, the breakeven condition is:

$$0.126L - 3.21E + 0.07\,(CRF) + 0.9 = 0$$

or

$$L = 25.5E - 0.56\,(CRF) - 7.1$$

This equation is shown in Figure 7-2 for CRF's of 10% and 15%.

It is clear that the electrolytic option LHTM cannot be competitive with the LTM option at power rates of 2¢/kWhr.

Power rates of 1¢/kWhr would have to be negotiated to make the LHTM option attractive. The LHTM option also incurs a penalty for higher lignite prices. The LHTM option is, therefore, abandoned.

7.3 THE LTM VERSUS THE GTM* OPTION

The question here is to find out whether methanol produced in Onakawana using the LTM option, and shipped[1] to southern Ontario, can be competitive with methanol produced in southern Ontario using natural gas.

Based on unit train loads, the difference in transportation costs between Onakawana - Toronto and southern Ontario - Toronto is about 1.9¢/litre. This difference must be added to the LTM operating fixed cost of 3.7¢/litre. The cost equations of the LTM and GTM* options are:

$$\text{LTM} \quad M(¢/l) = 0.315L + 0.88\,(CRF) + (3.7 + 1.9)$$
$$\text{GTM*} \quad M(¢/l) = 2.63G + 0.30\,(CRF) + 2.0 \text{ (from section 4.6)}$$

The breakeven condition is:

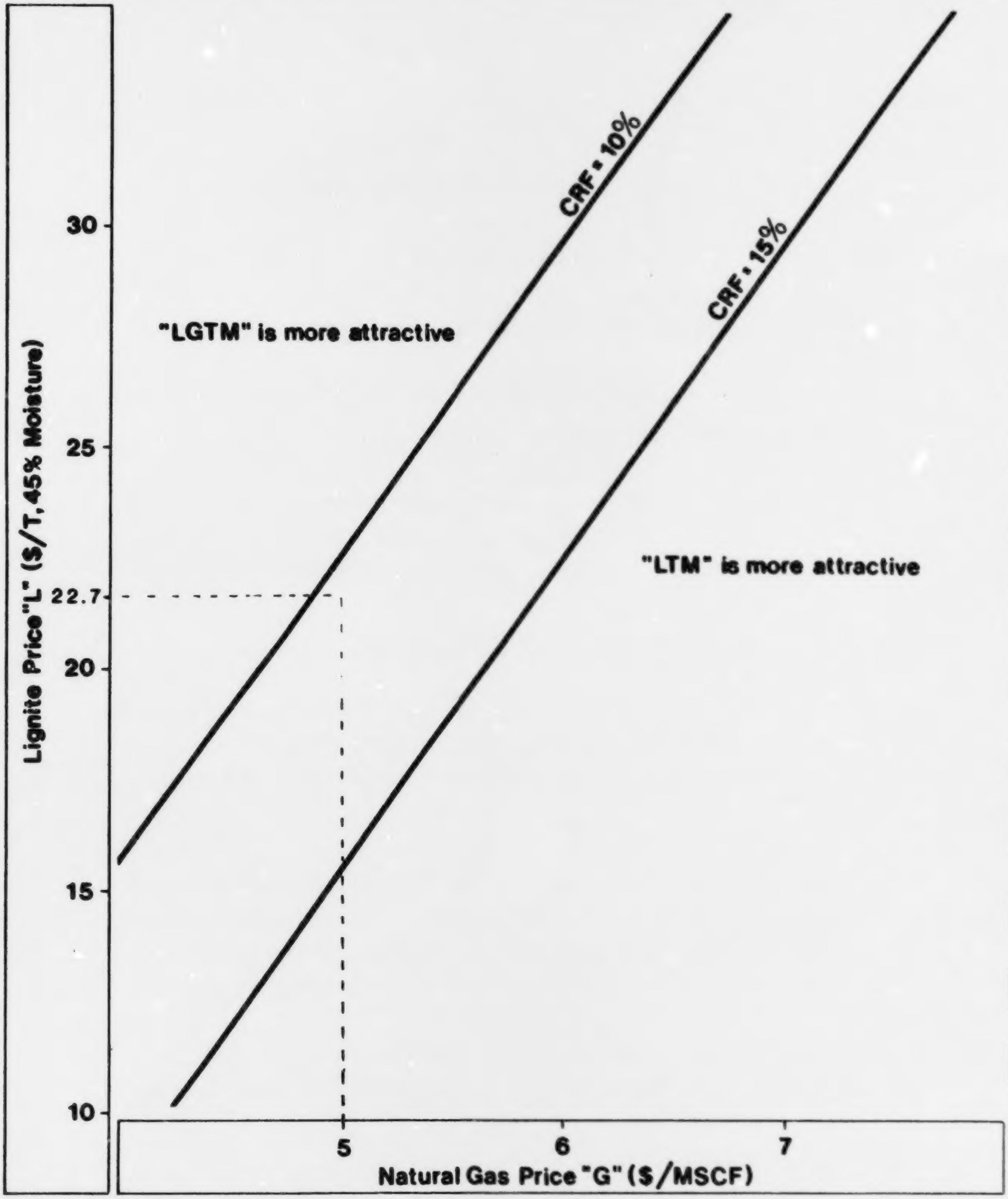
$$0.315L - 2.63G + 0.58\,(CRF) + 3.6 = 0$$

or

$$\boxed{L = 8.35G - 1.84\,(CRF) - 11.4}$$

---

[1] Transportation costs have been developed in a separate study.



Comparison of the "LTM" and "LGTM" Options
Figure 7-1

Figure 7-3 shows that if lignite is priced at \$16/T, the LTM option is competitive with the GTM* option when gas prices rise to \$5.5-\$6.6. They could reach this level by the mid 1990's. To narrow the time frame, we looked at the probable methanol plant gate selling prices developed by OEC in a separate report.

In that analysis, estimates were made of the methanol price that could be paid at the plant gate if neat methanol was retailed at the same price as gasoline after adjusting for differences in volumetric energy content. The resulting southern Ontario plant gate constant dollar prices are 18.1¢/litre, 21.5¢/litre and 22.5¢/litre for the years 1990, 1995 and 2000.

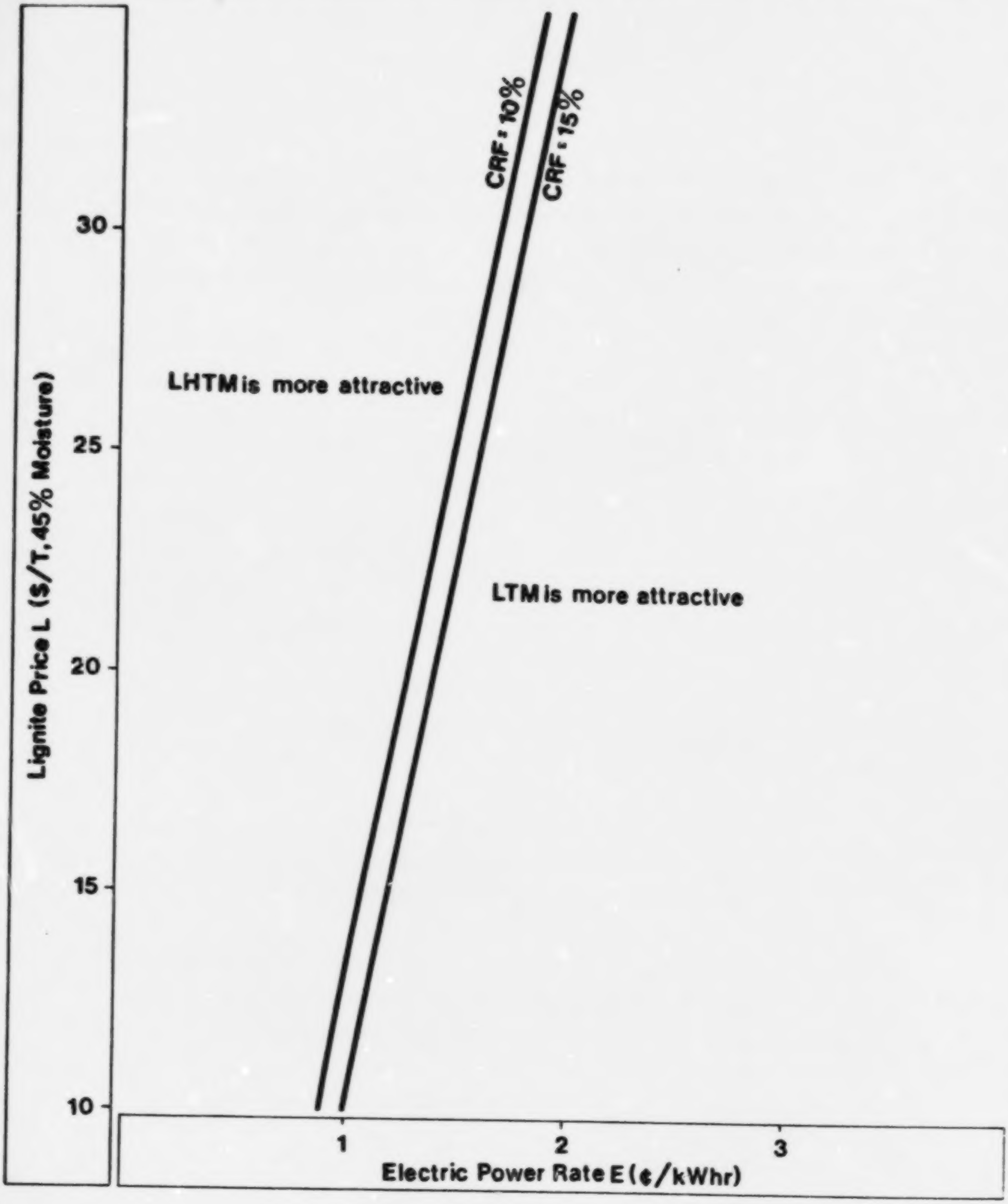
Ministry of Energy projections suggest natural gas prices in the order of \$5.34, \$6.22 and \$6.69/MSCF in 1990, 1995 and 2000, respectively.

Using the GTM* equation and applying the methanol selling price to M, and the projected gas price to G, we can determine the capital recovery factor:

1990 $18.1 = (2.63 \times 5.34) + 0.30\,(CRF) + 2.0$
1995 $21.5 = (2.63 \times 6.22) + 0.30\,(CRF) + 2.0$
2000 $22.5 = (2.63 \times 6.69) + 0.30\,(CRF) + 2.0$

Solving for CRF, we obtain values of 7% in 1990 and about 10% in 1995 and 2000. When the realization price M is equal to the actual selling price, as is the case in this calculation, the CRF value equals the return on investment (ROI). Therefore, estimated methanol selling prices in southern Ontario, during the 1990's, can produce a 10% maximum real return for the GTM* option.

Referring to Figure 7-3 and considering only the 10% CRF line, the LTM option looks more attractive. Assuming lignite prices remain in the \$15 - \$17 range, the LTM option can be considered as a viable alternative as soon as gas prices break the \$5.5/MSCF mark, perhaps by 1991. This also assumes that methanol demand justifies a large scale plant.



Breakeven of the LTM and LHTM Options
Figure 7-2

Figure 7-3 shows that if lignite is priced at \$16/T, the LTM option is competitive with the GTM* option when gas prices rise to \$5.5-\$6.6. They could reach this level by the mid 1990's. To narrow the time frame, we looked at the probable methanol plant gate selling prices developed by OEC in a separate report.

In that analysis, estimates were made of the methanol price that could be paid at the plant gate if neat methanol was retailed at the same price as gasoline after adjusting for differences in volumetric energy content. The resulting southern Ontario plant gate constant dollar prices are 18.1¢/litre, 21.5¢/litre and 22.5¢/litre for the years 1990, 1995 and 2000.

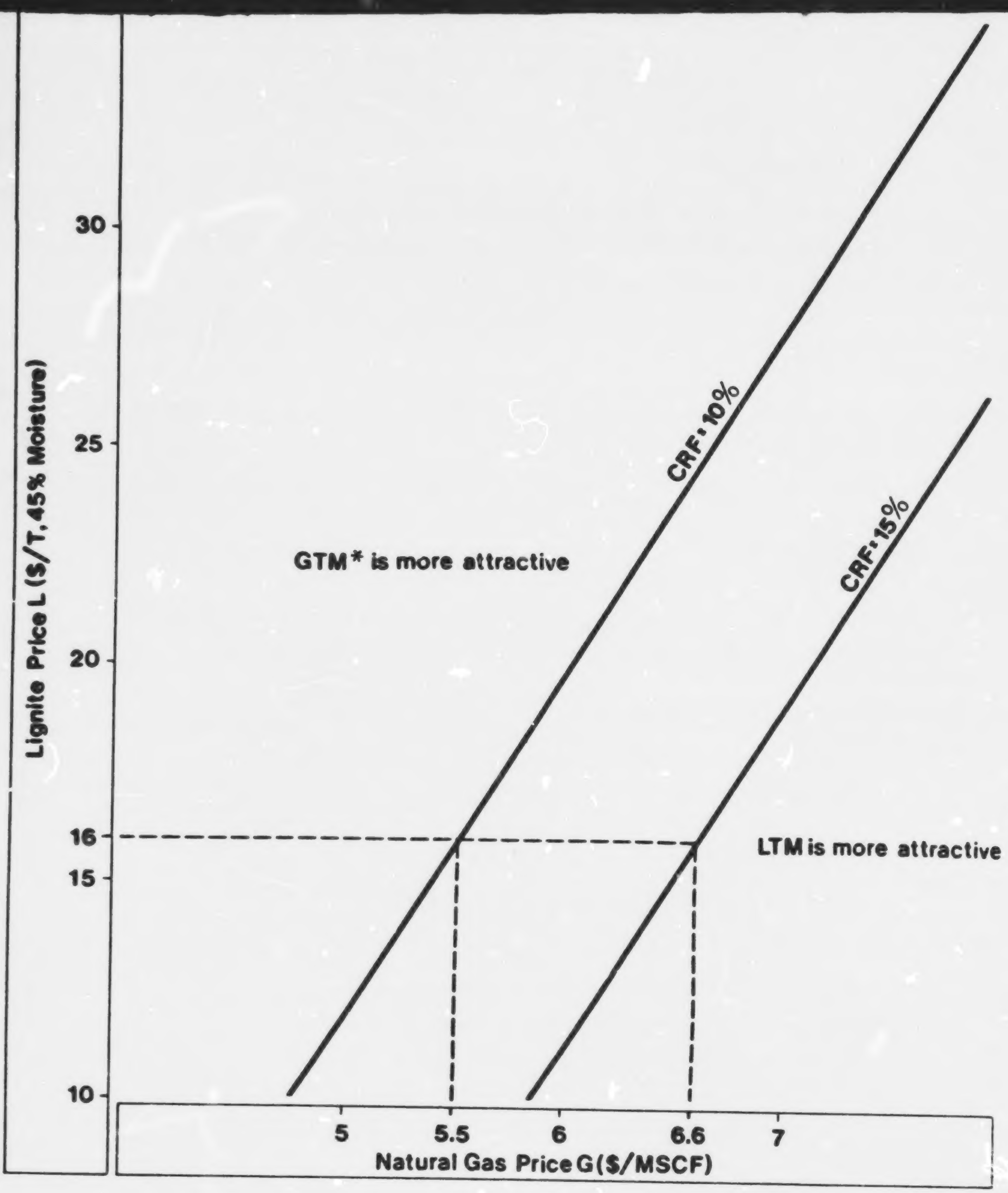
Ministry of Energy projections suggest natural gas prices in the order of \$5.34, \$6.22 and \$6.69/MSCF in 1990, 1995 and 2000, respectively.

Using the GTM* equation and applying the methanol selling price to M, and the projected gas price to G, we can determine the capital recovery factor:

$$\begin{array}{llcl}
1990 & 18.1 & = & (2.63 \times 5.34) + 0.30\,(\mathrm{CRF}) + 2.0 \\
1995 & 21.5 & = & (2.63 \times 6.22) + 0.30\,(\mathrm{CRF}) + 2.0 \\
2000 & 22.5 & = & (2.63 \times 6.69) + 0.30\,(\mathrm{CRF}) + 2.0
\end{array}$$

Solving for CRF, we obtain values of 7% in 1990 and about 10% in 1995 and 2000. When the realization price M is equal to the actual selling price, as is the case in this calculation, the CRF value equals the return on investment (ROI). Therefore, estimated methanol selling prices in southern Ontario, during the 1990's, can produce a 10% maximum real return for the GTM* option.

Referring to Figure 7-3 and considering only the 10% CRF line, the LTM option looks more attractive. Assuming lignite prices remain in the \$15 - \$17 range, the LTM option can be considered as a viable alternative as soon as gas prices break the \$5.5/MSCF mark, perhaps by 1991. This also assumes that methanol demand justifies a large scale plant.



Breakeven of the LTM and GTM* Options
Basis: Methanol shipped from
Onakawana to Southern Ontario
Figure 7-3

The comparison of the GTM* and LTM option is shown in Table 7-1. A point worth considering here is the impact of different market prices. If methanol prices are relatively low, as shown in the table, the expected return on investment is correspondingly low and the capital intensive LTM alternative is favoured. If selling prices were significantly higher than the ones used, for example 3¢/litre more, then the ROI approaches 20% for GTM* and the LTM option loses its viability for the rest of this century.

However, based on OEC's investigations, it appears that realizing an ROI significantly higher than 10% is rather unlikely. The LTM option is a serious alternative to the GTM* conventional route.

Table 7-1

EVALUATION OF THE LTM OPTION

	1990	1995	2000	Remarks
Gas Price G ($/MSCF)	5.34	6.22	6.69	Average of MEny Projections (See Figure 6-1)
Methanol plant gate price (¢/l)				
in Southern Ontario	18.1	21.5	22.5	Allowing for shipping costs to southern Ontario
in Onakawana	16.2	19.6	20.6	
Lignite price ($/T)	16.2	16.6	17.0	Assuming lignite at $16.2/T in 1990, escalating at 0.5% per year
GTM* Option				
CRF (%)	6.9	10.5	9.7	
LTM Option				
CRF (%)	8.4	12.1	13.1	

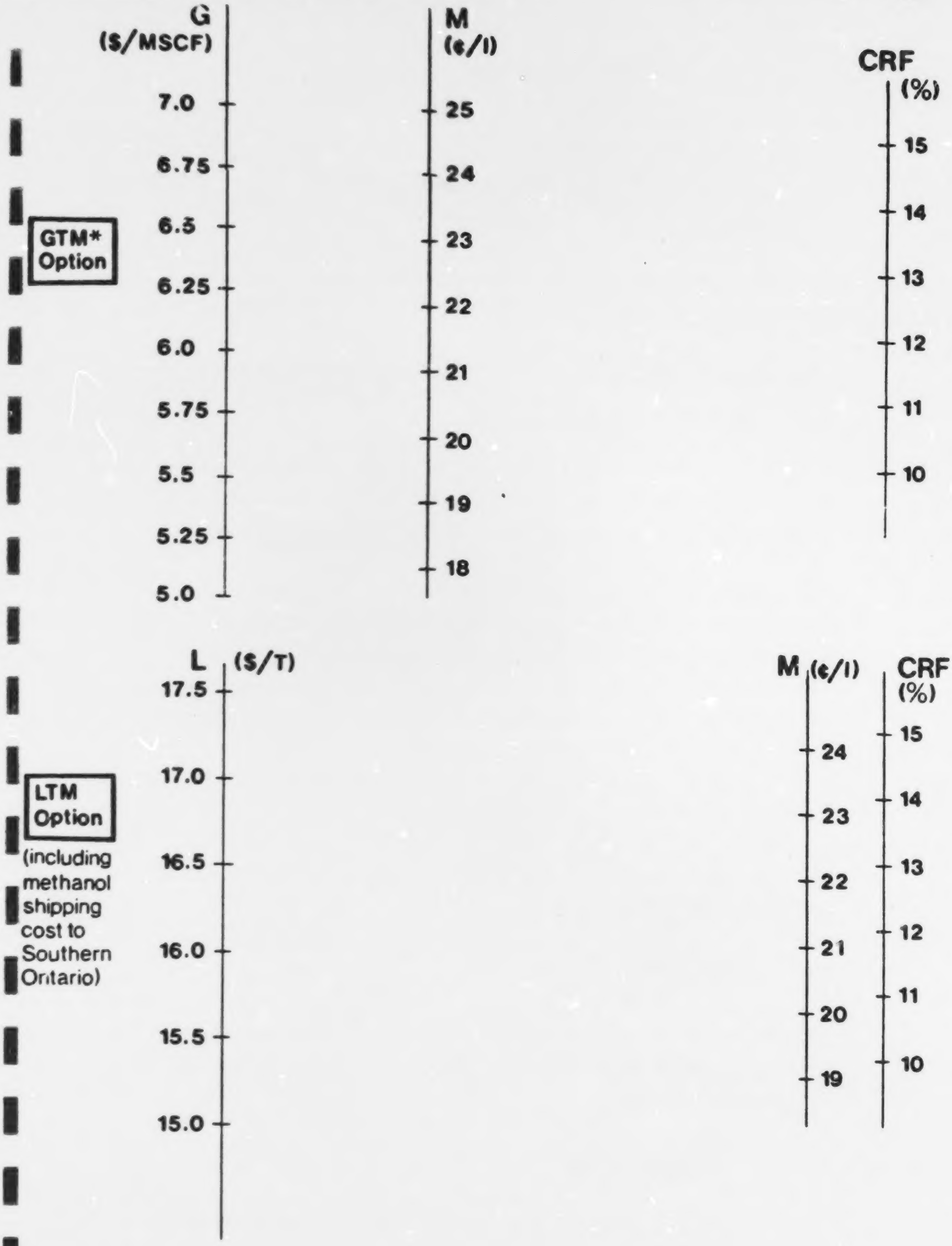
7.4 SENSITIVITY ANALYSIS

The series of equations presented in sections 4.6 and 5.6.6 and the breakeven curves shown in Figure 7-1 through 7-3 provide the answers to "what if" questions regarding feedstock cost levels and capital recovery factors.

The actual realization prices of methanol produced by the LTM and GTM* options are in Figure 7-4. The two nomographs relate the sensitivity of methanol cost to combinations of feedstock price and capital recovery factors.

6.5 PRELIMINARY CONCLUSIONS

Only the LTM lignite option is carried forward to the discounted cash flow analysis in Chapter 8. The GTM* is also carried forward as a reference case.



Methanol Realization Price in Southern Ontario (2500 TPD Plant Capacity)
Figure 7-4

8– DISCOUNTED CASH FLOWS

Two options are analyzed in this chapter: The large scale LTM and its corresponding GTM* option (2500 TPD).

The basis for the discounted cash flow (DCF) analysis is described in Chapter 3.

First, we define a base case for each option, by assuming probable values of prices, costs and plant operating data. The internal rates of return of the base case $(IRR)_0$ are calculated, in real terms, for the five options.

Next, we analyze nine sensitivity cases by modifying some of the most critical values of the base case such as feedstock costs and methanol price. The new values of the IRR are then compared to $(IRR)_0$. This enables us to identify the economic impact of these changes and to select the most potentially attractive production option.

Finally, we evaluate the effect of timing on the commercial viability of the project by delaying plant start-up by five years.

The computer print-outs of the base case and the sensitivity cases are presented in Appendix 3.

8.1 BASE CASE

All dollar values are constant 4Q 1983 Cdn $. All the data used in the base case is summarized in Table 8-1. We based our calculations on the following assumptions.

8.1.1 Plant Start-up Date

The year of plant start-up was assumed to be 1990, although we recognize that the demand for fuel methanol may not justify the construction of a plant as early as 1990.

TABLE 8-1

DCF BASE CASE ASSUMPTIONS

	LTM	GTM*	REMARKS
TPC	877	287	in millions $
Working Capital	12	23	(4Q 1983 Cdn $)
Total Investment	889	310	
Construction Years	4	3	
Capital Distribution Ratios			
Year 1986	0.10	-	Fraction of the
Year 1987	0.30	0.20	total plant cost
Year 1988	0.35	0.40	spent during
Year 1989	0.25	0.40	construction
SPDY	320	330	Days per year
Utilization Factors	0.75/ 0.9/1	0.8/1	
Methanol Price (¢/l)	17.6	19.2	Starting prices
Lignite Price (%$/T)	16.2	-	in 1990 (first
Natural Gas Price ($/MSCF)	-	5.34	operating year)
Lignite Use (T/litre)	0.00315	-	Refer to
Gas Use (MSCF/litre)	-	0.0263	linear equations
Operating Costs			
Manpower	13.1	5.5	Initial Value
Maintenance	18.6	11.8	for 1990
Utilities	3.8	1.3	(MM$/Yr)
Tax & Insurance	6.1	4.6	(4Q 1983 $)
Escalation Factors Applied to:			
Lignite Price	1.005	-	These escalation
Gas Price	-	1.0227	rates are shown
Methanol Price	1.016	1.016	as annual
Manpower Costs	1.020	1.020	multiplying factors
Maintenance Costs	1.020	1.020	
Utilities Costs	1.010	1.010	
Tax & Insurance Costs	1.005	1.005	

Feedstock and operating costs, which are likely to escalate in real terms from their estimated current value, were increased accordingly.

8.1.2 Total Investment

Total plant costs (TPC) and working capital are as calculated in Chapters 4 and 5. The TPC was not escalated to 1990 because the plant cost index follows closely the average rate of inflation.

Working capital is assumed to cover the total production costs (feedstock costs + operating costs) for a period of 45 days. The working capital was kept constant for each option since the impact of any change is considered minimal on the total investment. The total investment is the sum of the TPC and the working capital.

8.1.3 Construction Schedule

The lignite option (LTM) was based on a four year construction schedule, due to its larger size and its remote location.

To account for its relative simplicity in engineering and construction, the corresponding GTM* options was given a credit of one year in its construction schedule.

Option	Construction Schedule
LTM	4 years
GTM*	3 years

When the IRR is below 10%, in real terms, as is the case in most of these alternatives, an extension of one year in the construction schedule has a small effect on the overall return as illustrated in Table 3-1.

8.1.4 Projected Capital Outlays

The total plant cost (TPC) is distributed over the construction period in a way that approximates the expected capital expenditures. Except for long delivery items, actual expenditures are generally small at the early stages of a project. These expenditures rise dramatically in the second year when field work starts and, later, when major equipment is installed.

We used these capital distribution ratios.

Number of Years of Construction	Capital Distribution Ratios
3 years	0.2, 0.4, 0.4
4 years	0.1, 0.3, 0.35, 0.25

Working capital is added to the expenses incurred in the year before start-up (last construction year).

8.1.5 On-Stream Days per Year (SDPY)

We use the same values as in Chapters 4 and 5.

- 320 DPY for the LTM option
- 330 DPY for the GTM* option

The SDPY's refer to the total number of equivalent days per year during which the plant operates at design capacity.

For example, the LTM option is based on a design capacity of 2500 TPD. This plant is expected to run 320 days per year at design capacity, producing (320 x 2500) = 800,000 Tonnes per year. An allowance of 45 days per year (365 - 320) is included for planned and unscheduled shutdowns.

These SDPY's were estimated on the basis of plant scale, spare units and the development status of the gasification technologies.

8.1.6 Plant Utilization Factors

The expected annual production, as defined by the SDPY's is generally not achieved in the first year(s) of operation. This is because personnel are not fully trained and technical problems may have been overlooked at the engineering stage.

We assumed the following plant utilization factors.

Option	First Year	Second Year	Subsequent Years
GTM*	0.8	1.0	1.0
LTM	0.75	0.9	1.0

For example, the LTM plant will only produce (0.75 x 800,000) = 600,000 Tonnes the first year, (0.9 x 800,000) = 720,000 Tonnes the second year, and 800,000 Tonnes thereafter.

8.1.7 Methanol Selling Price

The projected selling price, at the gate of a plant located in southern Ontario, was estimated at 18.1¢ in 1990, expressed in 1983 dollars.[1] This price was based on a very low annual demand which does not justify the construction of a methanol plant in Ontario.

[1] Table 7.1

By 1995, the selling price in southern Ontario is estimated to increase to 21.5¢/litre. To test production economics, we assumed that volumes would be sufficiently high to yield a plant gate price part way between the estimated 1990 and 1995 levels of 18.1¢ and 21.5¢. A factor of one third was used to derive a 1990 plant gate price of 19.2¢/litre. This was the price used as the starting plant gate price, in 1990, for the GTM* option, based in southern Ontario.

The methanol price applied at the gate of the Onakawana plant, for the LTM option was modified to include:

- penalty of 1.9¢/litre to account for methanol transportation to the southern Ontario markets;
- credit of 0.3¢/litre to account for the sulphur and ammonia by-products produced in the lignite gasification process.[2]

$$19.2 - 1.9 + 0.3 = 17.6¢/\text{litre}$$

8.1.8 Lignite Price

Lignite price projections obtained from Manalta Coal show a current price of $15.75/T and an estimated escalation rate of 0.5% per year.[3]

Projected 1990 Lignite Price

$$15.75 \times (1.005)^6 = \$16.2/\text{T} \text{ (at 45\% moisture)}$$

---

[2] assuming $100/T of sulphur and $240/T of ammonia

[3] Chapter 6

8.1.9 Natural Gas Prices

In the base case, we assume the average value of the low and high cases projected by the Ontario Ministry of Energy, as shown in Table 6-1.

Projected 1990 Natural Gas Price (average value)

$5.34/MSCF

8.1.10 Feedstock Consumption

The feedstock consumption per litre of methanol is shown in the realization price linear equations, presented in Chapters 4 and 5.

8.1.11 Operating Costs

Operating costs are broken down as shown in section 3.1.2. These items were estimated, under current conditions, in Chapters 4 and 5. The estimated operating costs for 1990 were calculated by using these escalation factors, expressed in real terms:

Manpower Costs	2% per year
Maintenance Costs	2% per year[5]
Utilities	1% per year
Taxes and Insurance	0.5% per year

For example, the operating costs of the LTM option, as shown in Table 5-5 were prorated as:

---

[5] including catalysts and chemicals

Operating Costs	Current Value (1983 $MM/Year)				Projected 1990 Value (1983 $MM/Year)
Manpower	11.6	x	$(1.02)^6$	=	13.1
Maintenance	16.5	x	$(1.02)^6$	=	18.6
Utilities	3.6	x	$(1.01)^6$	=	3.8
Tax & Insurance	5.9	x	$(1.005)^6$	=	6.1
Total	37.6				41.6

8.1.12 Escalation Rates

Annual escalation rates were assumed for feedstock prices, methanol selling price and operating costs, all expressed in real terms.

o Lignite Price	0.5%/year
o Natural Gas Price	2.27%/year

(Estimated from the average gas prices forecast in 1990 ($5.34/MSCF) and 2000 ($6.69/MSCF), the escalation rate is 2.27% per year.)

o Methanol Selling Price 1.6%/year

(Estimated from the projected plant gate prices of 19.2¢/litre in 1990 and 22.5¢/litre in 2000, the escalation rate is 1.6% per year.)

o Operating Costs

The escalation rates are shown in section 8.1.11.

8.2 DCF EVALUATION OF THE BASE CASE

The discounted cash flow calculations, as shown in the computer print-outs in Appendix 3, are based on:

- constant 4Q 1983 Cdn $
- 20 year production life time with no salvage value
- gross profits before tax
- discounting to the first year of project implementation
- all operating costs except utilities, considered as fixed costs, regardless of the plant throughput
- all IRR's in real terms, above the inflation rate

The base case internal rates of return $(IRR)_o$ are listed in the next table.

TABLE 8-2

BASE CASE INTERNAL RATE OF RETURN $(IRR)_o$

Options	LTM	GTM*
$(IRR)_o$	7.5	4.2

- The LTM option provides the best IRR. It shows a significant advantage over the GTM* route. A 7.5% IRR in real terms translates into a 15% IRR in actual terms, at 7% inflation rate. While this IRR is clearly unsatisfactory for a project of this kind, leveraged financing may actually result in a higher return on equity (ROE). This aspect is discussed in Chapter 9.

- A closer look at the computer print-outs shows that the GTM* option produces steadily decreasing cash flows. The natural gas feedstock escalates at a higher rate than the methanol selling price.

The opposite occurs for the LTM alternative where lignite prices are expected to escalate at a slower rate than the 1.6% assumed for methanol.

The relative escalation rates of feedstock and product prices are indicative of the long term viability of a project.

8.3 SENSITIVITY CASES

The internal rate of return was found particularly sensitive to methanol and feedstock prices. To a lesser degree, the IRR was affected by operating costs, capital costs and plant construction schedules.

Items tested for sensitivity are listed in order of importance.

- methanol price
- feedstock price and escalation rate
- operating costs
- total plant cost
- methanol price escalation rate
- on-stream days per year
- escalation of manpower and maintenance costs
- construction schedule

To assess net impact, only one item was changed at a time, leaving all the other values as in the base case.

8.3.1 CASE A Methanol Price

A variation of ±10% of the base case price was tested. Given the uncertainly of the methanol price structure, the escalation rate was kept unchanged, at 1.6% per year. Tests on the variation of the escalation rate were done separately in Case E.

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8.3.2 CASE B$_1$ Feedstock Price and Escalation Rates

LIGNITE Manalta Coal estimated the current price of lignite at $15.75 + 15%. This price is within the $13.39 - $18.11 range. Prorating these prices to 1990 we arrived at:

Low Case	o initial price	$13.8/T
	o escalation rate	0.4%/year
High Case	o initial price	$18.7/T
	o escalation rate	0.6%/year

A higher base price was assumed to increase the escalation rate and vice-versa.

NATURAL GAS Here we applied the high and low cases defined in the price projections of the Ontario Ministry of Energy. The escalation rates were calculated for each case, based on the projected prices for 1990 and 2000.

Low Case	o initial price	$4.91/MSCF
	o escalation rate	2.16%/year
High Case	o initial price	$5.77/MSCF
	o escalation rate	2.37%/year

8.3.3 CASE C Total Operating Costs

A ±20% variation was tested on the total operating costs for the initial values in 1990. The escalation rates of the various components - manpower, maintenance, utilities, tax and insurance - remained unchanged from the base case.

8.3.4 CASE D Total Plant Cost

A ±10% variation was applied to the TPC. The working capital was unchanged from the base case. For the LTM case, a +20% variation was also tested.

8.3.5 CASE E Methanol Price Escalation Rate

The sensitivity test was carried out independently of Case A where the base selling price of methanol was altered.

Here, the methanol initial price kept the same value as in the base case, but escalation rates were varied by ±10%.

Low Case	o escalation rate	1.45%/year
High Case	o escalation rate	1.76%/year

8.3.6 CASE F On-Stream Days Per Year

The SDPY's were changed by ±10 days per year for both cases to assess the impact of a change of about ±3% in plant throughput.

8.3.7 CASE G Escalation of Manpower and Maintenance Costs

In the base case, a conservative 2% escalation rate was assumed for these two items. Here we reduce this rate to 1%.

8.3.8 CASE H Construction Schedule

We increase the construction period by one year for all the options.

GTM*	4 years
LTM	5 years

8.4 RESULTS OF THE SENSITIVITY TESTS[6]

8.4.1 The LTM Option

This option always shows a positive IRR, regardless of the case tested. Compared to the GTM* option, it proves more attractive unless:

- natural gas prices escalate as predicted by MEny's low case projections for the next 25 years, or
- methanol selling price, at the plant gate, in southern Ontario is higher than 20.5¢/l in 1990.

An increase of methanol selling price by 10% in Case A has a greater impact on the IRR of the LTM option than an increase of 15% of the lignite price in Case B_1.

This seems to indicate that if the LTM plant could be built in southern Ontario, the costs incurred by lignite drying and shipping may be offset by increased revenues from higher methanol plant gate prices. A southern Ontario plant would also cost less to build and operate. This alternative is worth considering.

8.4.2 Cases F, G and H

These show only a small impact on the IRR. The general viability of any option is not significantly affected by:

- small variations in the plant throughput
- escalation rates of operating costs
- plant construction schedule

---

[6] Table 8-3

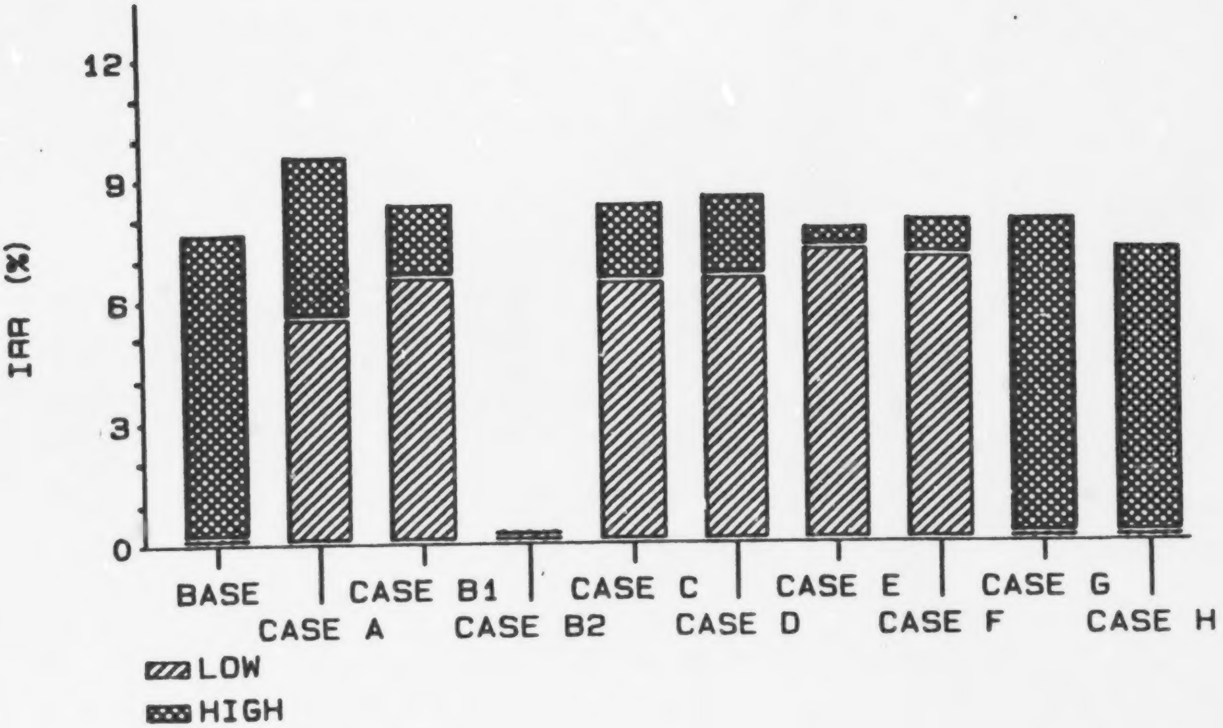
TABLE 8-3

DCF SENSITIVITY ANALYSIS

INTERNAL RATES OF RETURN (IRR)%

Sensitivity Cases	LTM	GTM*	Remarks
Base Case $(IRR)_o$	7.5	4.2	See Table 8-2
CASE A	5.5-9.5	Neg-12.4	Methanol Price
CASE B_l	6.5-8.3 -	- Neg-10.1	Variations of Lignite and Gas Prices
CASE C	6.4-8.3	1.3-6.2	Operating Costs (+20%)
CASE D	6.5-8.5 5.7	3.2-5.2 -	±10% on TPC ±20% on TPC (for LTM only)
CASE E	7.2-7.7	2.7-5.6	Methanol Escalation Rate
CASE F	7.0-7.9	3.5-4.9	SDPY (+10 days/year)
CASE G	7.8	5.0	Manpower and Maintenance Escalation Rate (1% per year)
CASE H	7.1	4.0	1 Year Added to Construction Schedule

These values are shown graphically in Figures 8-1 and 8-2



CASE A : High/low methanol selling price (+/- 10%)

CASE B1: High/low main feedstock price (+/- 10% for wood and peat)
(+/- 15% for lignite)
(+/- 8% for natural gas)

CASE B2: High/low electric power rate (+/- 10%)

CASE C : High/low operating costs (+/- 20%)

CASE D : High/low total plant cost (+/- 10%)

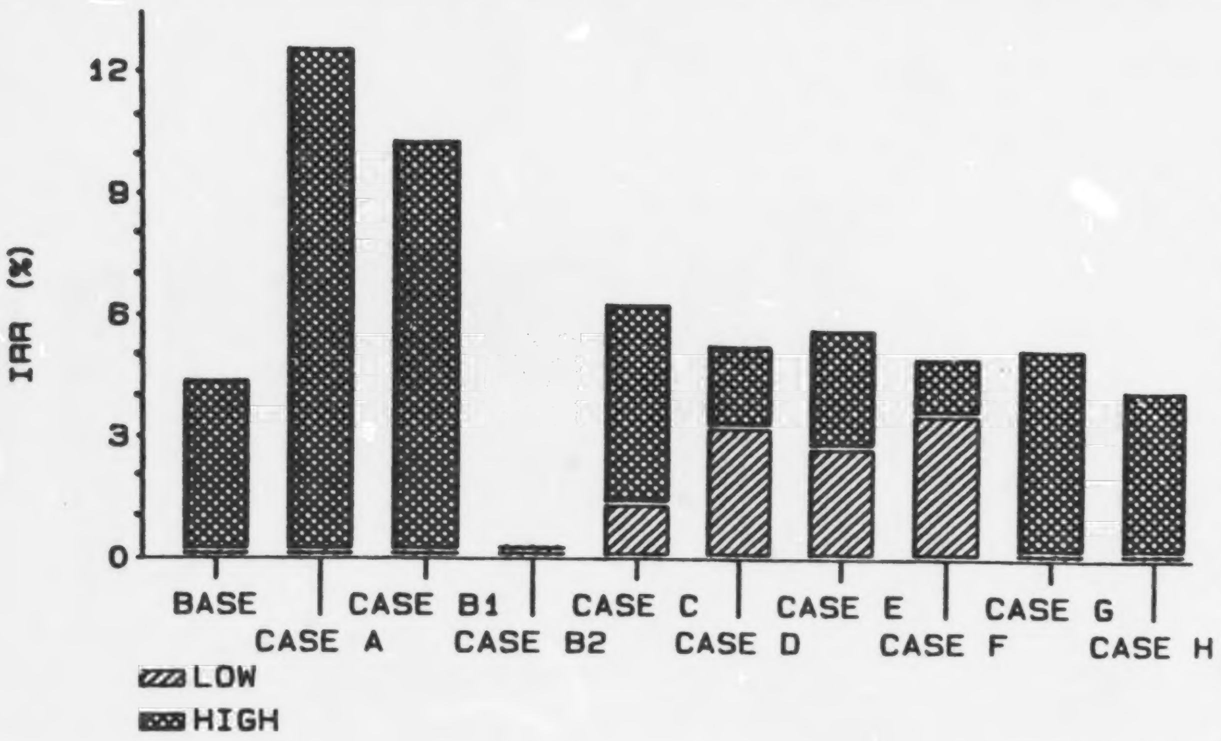
CASE E : High/low methanol price escalation rate (+/- 10%)

CASE F : High/low on-stream rate (+/- 10 days per year)

CASE G : Low manpower and maintenance sealation rate (1% per year)

CASE H : High plant construction schedule (+ 1 year)

Figure 8-1 DCF Sensitivity Analysis: LTM Process



CASE A : High/low methanol selling price (+/- 10%)

CASE B1: High/low main feedstock price (+/- 10% for wood and peat)
(+/- 15% for lignite)
(+/- 8% for natural gas)

CASE B2: High/low electric power rate (+/- 10%)

CASE C : High/low operating costs (+/- 20%)

CASE D : High/low total plant cost (+/- 10%)

CASE E : High/low methanol price escalation rate (+/- 10%)

CASE F : High/low on-stream rate (+/- 10 days per year)

CASE G : Low manpower and maintenance sealation rate (1% per year)

CASE H : High plant construction schedule (+ 1 year)

Figure 8-2 DCF Sensitivity Analysis: GTM* Process

8.5 PROSPECTS FOR COMMERCIAL VIABILITY

The internal rates of return listed in Table 8-3 were calculated on the assumption that both plants would start in 1990, anticipating a developed market for methanol.

Fuel methanol market forecasts are difficult to establish because of the number of preconditions that must be resolved.

A large scale LTM plant, producing about one billion litres per year, would not likely come on-stream before 1995.

In this new time perspective, the LTM and GTM* options show results significantly different from those listed in Table 8-3

Cases A and B_1 are particularly important because high methanol selling prices and low feedstock prices proved to be the most unfavourable when comparing the LTM option to the GTM* option.

If the large scale plants, LTM and GTM*, come on-stream in 1995 instead of 1990, the initial price of natural gas is projected to be about $0.9/MSCF higher. The lignite price will have increased only marginally.

This difference in the escalation rate of the feedstocks is reflected in the values of the new IRR's shown in Table 8-4.

TABLE 8-4

TIME EFFECT ON ECONOMIC VIABILITY

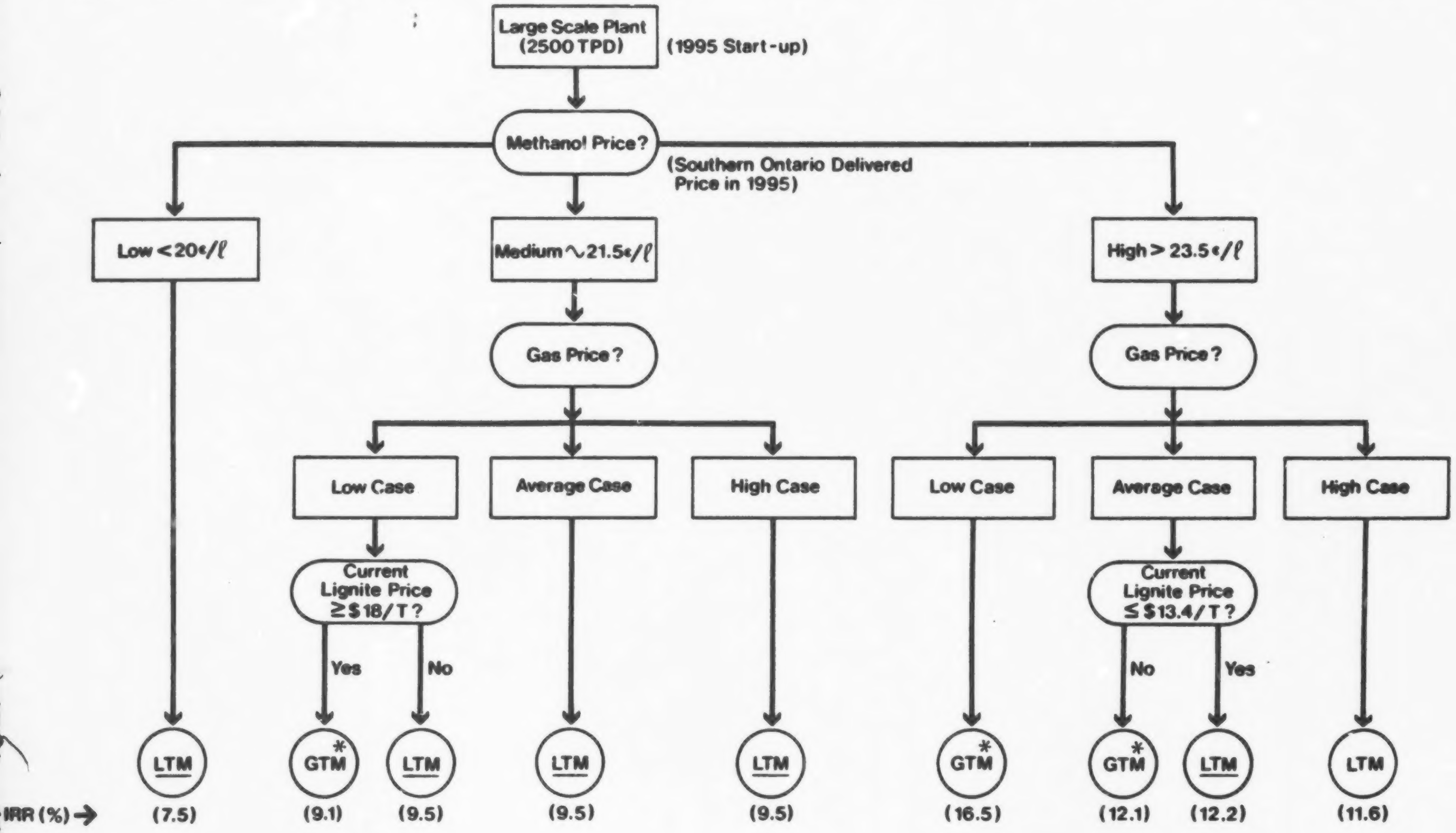
	INTERNAL RATE OF RETURN		
	LTM	GTM*	REMARKS
CASE A			
1990 Start-up	9.5	12.4	High Methanol Price Case Only
1995 Start-up	11.6	12.1	
CASE B_1			
1990 Start-up	8.3	10.1	Low Gas and Low Lignite Price Case Only
1995 Start-up	10.3	9.1	

The IRR of the GTM* option decreases when plant start-up is delayed, while the opposite occurs for the LTM option.

Therefore, if a large scale plant is projected for 1995, the LTM option becomes more attractive.

8.6 OVERALL EVALUATION

Conclusions are summarized in the logic diagram shown in Figure 8-3. Except for the combined case when methanol prices are high and, at the same time, gas prices are low, the alternative LTM process is consistently more attractive than the conventional GTM processes.


Large Scale Plant (2500 TPD)
(1995 Start-up)
Methanol Price?
(Southern Ontario Delivered Price in 1995)
Low < 20¢/ℓ
Medium ∿ 21.5¢/ℓ
High > 23.5¢/ℓ
Gas Price?
Gas Price?
Low Case
Average Case
High Case
Low Case
Average Case
High Case
Current Lignite Price ≥ $ 18/T?
Yes
No
Current Lignite Price ≤ $ 13.4/T?
No
Yes
LTM
GTM*
LTM
LTM
LTM
GTM*
GTM*
LTM
LTM
IRR (%) →
(7.5)
(9.1)
(9.5)
(9.5)
(9.5)
(16.5)
(12.1)
(12.2)
(11.6)

Alternatives for a Large Scale Plant
Figure 8-3

9 — PRELIMINARY FINANCIAL ANALYSIS

This chapter presents a preliminary financial analysis of the most promising option -- the lignite-to-methanol process, LTM. The purpose of this analysis is to assess the effects on the return on equity of typical fiscal and financing conditions which could apply to projects of this nature. However, the analysis is intended only as an illustration to augment the economic analysis in chapter 8. No attempt has been made to optimize the overall structure or to make use of all possible incentive programs for which the projects might be eligible.

9.1 PROJECTED OPERATING/ECONOMIC CONDITIONS

Based on the findings of chapter 8, we selected the timing, the operating conditions and the feedstock and product cost projections which appeared sufficient to make the project economically viable.

The assumptions applied to the LTM project are summarized in Table 9-1. Plant operating costs are as shown in Table 8-1, escalated to 1995 for the LTM project. All escalating factors are as shown on the DCF Base Case Assumptions.

Table 9-1 indicates an overall real internal rate of return before tax of 9.5% for the LTM project.

TABLE 9-1

PROJECTED OPERATING/ECONOMIC CONDITIONS (LTM PROJECT)

	LTM Project	REMARKS
- Project implementation date	1991	
- Years of construction	4	
- Plant start-up	1995	
- Total plant cost	877	- millions of dollars (1983$)
- Working Capital	12	- millions of dollars (1983$)
- Total investment	889	- millions of dollars (1983$)
- SDPY	320	- on-stream days per year
- Plant Utilization Factor	0.75/0.9/1.0	- Fraction of nominal capacity during first operating years
- Lignite price ($/T)	16.6	- In 1995 (1983 Dollars), assuming base case projections, as shown in Chapter 8
- Initial Methanol selling price (¢/l)	19.9	- In 1995 (1983 Dollars), shipped to southern Ontario, and including by-product credits
- Internal rate of return IRR (%)	9.5	- In real terms. (see computer printouts in Appendix 3)

9.2 FINANCIAL ANALYSIS

The financial analysis assumes a conventional project form of organization and financial structure. The advantage of leverage is introduced through the funding of 75% of project capital requirements with debt securities. The balance is provided in the form of equity capital.

Project finance is a technique whereby lenders rely principally upon the cash generated by the project, rather than the resources of sponsors, to meet debt payment obligations. Project finance structures are frequently complex due to the necessity to allocate construction and operating risk among sponsors, lenders and users. In this evaluation, no attempt has been made to examine these issues or to develop the specifics of security arrangements that would be required to support a project financing along the lines illustrated.

9.2.1 Assumptions Used

- Inflation is assumed to be 7% per year.
- The project will be financed with 25% equity and 75% debt. Full equity contributions are received before debt obligations are undertaken. The structure of the long-term debt is:

Year Incurred	Principal (Millions)	Term	Year Principal Repayment To Start	Interest Rate
1992	$243	15 years	1997	13.5%
1993	$605	14 years	1997	13.5%
1994	$486	13 years	1997	13.5%

Principal repayments are assumed to be made evenly throughout each of the years. Principal is to be repaid over 10 years, with an equal amount repaid each year.

- A bank line of credit at 13.5% is assumed to cover cash deficiencies throughout the year.
- Interest on long term debt and the bank line has been capitalized during the period of construction (1991-1994).
- No depreciation or CCA is taken until the first year of operation, 1995.
- Depreciation has been calculated on the declining balance basis using the following rates:

Equipment	10%
Building	5%

- Capital cost allowance (CCA) has been calculated assuming the assets are
  in the following classes:

Equipment Class	29
Building Class	3

Scientific research tax credits and allowances and investment tax credits have not been accounted for in the projections.

- The corporate tax rate is assumed to 45%. Current tax law is assumed to remain in effect throughout the projected period. Maximum CCA has been taken even though the Company is in a loss position because utilization of the loss carry forward in future years is assumed to be virtually certain.
  The company is assumed to follow the deferral method of tax allocation to account for income taxes; deferred income taxes are based on timing differences between accounting income and income for tax purposes. For this project, the only timing difference assumed is between depreciation and capital cost allowance.
- Revenues and expenses are assumed to be collected and paid in the month to which they relate.

9.2.2 Results

Computer print-outs are included in Appendix 3. They show:

- the debt and capital disbursements, in nominal dollars, together with the principal and interest repayment schedules;
- the calculation of depreciation, capital cost allowance (CCA) and corporate tax;
- the projected income statement;
- the projected cash flows.

The estimated after tax returns on equity is 15.3%. This return is marginal if considered from a conventional financing perspective using typical industrial criteria. However, the LTM process involves new technology and is at a relatively early development stage. This project would not be contemplated before the 1990's and there is the possibility that the economics could be improved with further technical advances and optimization. As a result, returns of this level could be sufficient to attract initial investor interest, particularly if investors have a more direct interest in the project -- for example, a market use for the methanol. Investors would also have to be in a position to provide the financing support arrangements that lenders would demand.

In addition, the return on equity could be substantially improved if the analysis was refined to take account of tax credits, industrial development assistance programs or other government support mechanisms that might be available for projects of this nature. This project would involve continuing research and development activities and could make significant contributions to government energy, technology development, employment and industrial development objectives.

The project form of financing was selected for this analysis because it demonstrates the financial and economic characteristics of the process on a stand-alone basis. However, a complete financial evaluation would also have to consider partnership or joint venture types of organization and financing. These alternative structures could also result in improved returns to equity participants because of the potential to flow tax benefits back to the participants to be offset against other income.

On the basis of the calculated returns, the lignite to methanol project appears to be more attractive than smaller scale methanol plants based on wood or peat. However, the total investment is several times higher. This substantial difference in investment cost would undoubtedly have a significant bearing on investors decisions.

From a market point of view, smaller scale projects, based on wood or peat as main feedstocks, are expected to be implemented earlier. But, in the event that annual methanol demand in the early 1990's approaches the one billion litre mark, the lignite based project is unquestionably more attractive.

In general, the financial analysis supports the conclusions of the earlier technical and economic analysis. Taking into account the nature of the LTM project and its probable timing and stages of development, it appears promising from a private investment perspective, and should be considered for further development and/or demonstration.

ABBREVIATIONS

ATM	Atmosphere (s)
BTU	British thermal unit
CANMET	Canada Centre for Mineral & Energy Technology
CC	Carbon conversion
CRF	Capital recovery factor
DCF	Discounted cash flow
DCFROI	Discounted cash flow return on investment
GTM	Gas-to-methanol
HHV	Higher heating value
HTW	High Temperature Winkler
ICI	Imperial Chemical Industries Limited
IRR	Internal rate of return
KT	Koppers-Totzek
LGTM	Lignite and natural gas to methanol
LHTM	Lignite and electrolytic hydrogen to methanol
LHV	Lower heating value
LTM	Lignite to methanol
MEny	Ontario Ministry of Energy
maf	Moisture and ash free
mf	Moisture free
NPV	Net present value
OEC	Ontario Energy Corporation
PSA	Pressure swing absorption
psi	Pounds per square inch

ROE	Return on equity
ROI	Return on investment
SCF	Standard cubic foot (60°F)
SCFH	Standard cubic feet per hour
SDPY	On-stream days per year
T	Metric Tonne 1000 kg (2,205 lb)
TE	Thermal efficiency
TPC	Total plant cost
TPD	Tonnes per day
TPH	Tonnes per hour
TPY	Tonnes per year

SYMBOLS

CH_3OH/MEOH	Methanol
CH_4	Methane
CO	Carbon monoxide
CO_2	Carbon dioxide
G	Natural gas
Gcal	Gigacalories (10^6 kilocalories)
H_2	Hydrogen
H_2O	Steam/water
H_2S	Hydrogen sulphide
GJ	Gigajoules (= 0.24 Gcal)
Kcal	Kilocalories (3.97 BTU)
KPa	Kilopascals (1 psi = 6.89 kPa)
kWhr	Kilowatt-hour
L	Lignite
1 litre	(0.001 cubic metre)
M	Methanol
MSCF	Thousand standard cubic feet (60°F)
MW	Megawatt
MWhr	Megawatt-hour

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APPENDIX 1

ONTARIO LIGNITE INDIRECT LIQUEFACTION STUDY

By

KILBORN LTD.

Ontario Lignite
Indirect Liquefaction Study
Phase 0
Process Technology Assessment

Prepared for:
Ontario Energy Corporation

Kilborn Limited
April 1983

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Introduction

1.0 INTRODUCTION

Conversion of lignite to methanol or liquid transportation fuels using advanced lignite gasification/synthesis gas conversion technology is one of the best prospects for increasing conventional domestic energy output within the next decade. With vast reserves of lignite in Ontario, conversion to methanol or conventional liquid transportation fuels is of particular interest.

To evaluate the potential for lignite gasification/synthesis gas conversion in Ontario, Kilborn Limited was retained by Ontario Energy Corporation in February 1983. Management of the program is the responsibility of Ontario Energy Corporation and an Advisory Committee consisting of representatives from the following organizations:

- Ontario Energy Corporation
- CANMET, Energy Mines and Resources
- Onakawana Developments Limited
- Saskatchewan Oil & Gas Corporation

The objective of the Kilborn assignment is to "Determine the Best Process(es) for Converting Ontario Lignite including Onakawana Lignite into Methanol and to Establish the Corresponding Cost Data". The program is being carried out in four phases, the specific objectives of each phase being as follows:

Phase 0 - An assessment of four preselected gasification process technologies in terms of their capabilities to indirectly liquefy Ontario lignite including Onakawana lignite into methanol and other liquid fuels.

Phase 1 - Determine the gasification/conversion process(es) most economically attractive for the treatment of Ontario lignite including Onakawana lignite.

Phase 2 - Perform test program(s) on Ontario lignite including Onakawana lignite at the technology developer facility(ies) to

a) confirm the suitability of the technology(ies) to convert Ontario lignites to methanol.

b) provide process data for the selection of one single technology for preliminary engineering design in Phase 3.

Phase 3 - Provide preliminary engineering design and cost estimates for all process, offsite and utility systems based on the most attractive technology selected in Phase 2.

Although, as indicated earlier, four gasification process technologies were to be preselected by Ontario Energy Corporation and its Advisory Committee, only three were preselected, as follows:

- Koppers-Totzek Entrained Bed Process
- Westinghouse Fluidized Bed Process
- High Temperature Winkler Fluidized Bed Process

The fourth gasification technology to be investigated in Phase 0, a choice between the Lurgi Dry Ash Fixed Bed Process and the Texaco Coal Gasification Process (an entrained bed technology), was left to Kilborn. Based on a preliminary assessment of these two processes, Letter KIL/OEC-3 dated March 18, 1983, the Lurgi Process was recommended as the fourth technology to be investigated.

This report presents the results of the Phase 0 phase of the study. It contains all of the information and data used and referred to during the completion of this phase of the work and contains recommendations on which the gasification technologies should be carried forward into Phase 1 of the program.

KILBORN

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Lignite Indirect Liquefaction
O.E.C. Project LTM-01
Kilborn Project No. 3424-15
Letter No. KIL/OEC-3

March 18, 1983

Ontario Energy Corporation
101 Bloor St. West
5th Floor
Toronto, Ontario M5S 1P7

Attention: Mr. S. Benzimra
Technical Manager

Dear Mr. Benzimra:

Please find enclosed a brief report giving a justification of Kilborn's recommendation to evaluate the LURGI gasification technology rather than the TEXACO technology.

Yours very truly,

J. McMahon, P.Eng.
Study Manager

Encl.

cc: N.E. Anderson
C. Chao

JMcM/wn

COMPARISON OF THE
LURGI FIXED BED GASIFICATION PROCESS
AND THE
TEXACO COAL GASIFICATION PROCESS

Following are brief descriptions of the Lurgi Fixed Bed Gasification Process and the Texaco Coal Gasification Process along with a preliminary assessment of their ability to efficiently process high moisture content Ontario lignites. The results of this assessment were used to select the Lurgi Fixed Bed Gasification Process as the fourth process to be evaluated in the Ontario Energy Corporation Lignite Indirect Liquefaction Study.

Lurgi Fixed Bed Gasification Process

The Lurgi Gasification Process is a first generation pressurized, fixed-bed dry bottom gasification process, originally developed in the early 1930's as a low pressure, air blown system. Through continued development, the current version of the process operates at high pressure, can utilize either oxygen or air as oxidant and can efficiently produce a wide range of coals and lignites (most coals except swelling or caking type coals).

The Lurgi gasifiers are commercially proven as they have been used in the production of synthesis gas from coal at SASOL in South Africa for the past 25 years. Based on their commercial success and their ability to process high moisture content lignites, Lurgi gasifiers have been selected for the gasification of North Dakota lignite in the Great Plains Gasification Process.

Even though the Lurgi Gasification Process is commercially proven and can efficiently process high moisture content lignites, it does have several drawbacks. As the operating temperature varies throughout the fixed lignite bed, to a maximum of about 1000°C at the bottom of the bed, devolatilization and decarbonization of the lignite in the upper part of the bed results in the production of a raw synthesis gas containing condensible materials including oxygenated compounds and tars. In addition, as the temperature at the bottom of the bed must be maintained below 1000°C to prevent ash softening and agglomeration, a high ratio of steam-to-oxygen is required. As much of the steam passes through the bed, heat losses are increased and expensive treatment of large quantities of condensed foul water (containing phenols, ammonia and sulphur compounds) must be provided.

These drawbacks can be overcome through the appropriate selection of downstream processing units such that the Lurgi Gasification Process can be considered suitable for the production of high quality synthesis gas from high moisture content lignites.

Texaco Coal Gasification Process

The Texaco Coal Gasification Process is a second stage process incorporating a single stage, slagging, pressurized, entrained bed, down flow gasifier. The process has been adapted from the Texaco Partial Oxidation Process for the gasification of residual fuel oils in the presence of a mixture of oxygen and steam. Whereas most entrained bed gasifiers utilize a dry coal feed system and steam injection as moderator, the Texaco gasifier utilizes a concentrated coal/water slurry feed system. As such, the water contained in the slurry acts as a moderator.

As all materials leaving the gasifier are exposed to relatively high temperatures, in the order of 1200-1500^{o}C, virtually no by-products (phenols, tars, oils or condensible hydrocarbons) are formed and the slag is essentially inert. The raw synthesis gas contains hydrogen, carbon monoxide, carbon dioxide, water vapour and sulphur compounds.

Although the Texaco Coal Gasification Process has been proven successfully in demonstration plants and will be commercialized in the near future at the Tennessee Eastman chemicals plant in Tennessee, the process does not appear suitable for high moisture content Ontario lignites. As the total water content of the lignite/water slurry feed acts as a moderator to control gasification temperature, too high a water content will result in higher oxygen requirements and lower thermal efficiency to maintain the gasifier temperature of 1200-1500^{o}C, as determined by the lignite ash fusion or softening temperature.

For efficient operation total water content of the lignite/water slurry must not exceed 45-50 wt percent. As the run-of-mine Ontario lignites contain 40-55 wt percent moisture, it does not appear that efficient slurrying for a Texaco gasifier operation can be accomplished without deep predrying of the lignite (to 2-5 wt percent moisture). This is an expensive and commercially unproven process. On the basis of this apparent problem, the Texaco Coal Gasification Process is not recommended for inclusion in this study.

Conclusions and Recommendations

2.0 CONCLUSIONS AND RECOMMENDATIONS

The results of the assessment of the four selected gasification process technologies in terms of their potential to gasify Ontario lignites for the production of methanol and other liquid fuels are highlighted in the following sections of this Phase 0 report.

Key gasification process characteristics are summarized in Table 2.1 for each technology and are discussed in greater detail in Section 3.0 of this report. These characteristics of the various processes, along with the available physical, chemical petrographic and ash analyses for Onakawana, Ontario, Saskatchewan Coronach and reference lignites, were used as the basis for the conclusions and recommendations which follow.

2.1 CONCLUSIONS

- Based on the limited analyses available, the Onakawana and Ontario lignites appear to be suitable gasification process feedstocks.
- Ash content, sodium content of the ash and ash fusion temperatures of the Onakawana, Ontario and Saskatchewan Coronach lignites appear to be similar. As these are key gasification parameters, gasification performance is expected to be similar.
- Moisture content of the Ontario and Onakawana lignites appears to be 5-10 percent higher than for the Coronach lignite. This should not adversely affect gasifier performance but will increase predrying requirements and possible decrepitation of the lignite during the predrying operation.
- Each of the process developers has commercial or pilot-scale experience with lignites. Based on preliminary contacts, each developer indicates that his process should be expected to process Ontario lignites efficiently.

- Lignite predrying is required by all gasification processes, more extensive for the Koppers-Totzek, HT Winkler and Westinghouse processes than for the Lurgi process. Lurgi expressed a concern that lignite predrying may result in severe decrepitation and the generation of significant fines which cannot be processed in the dry-bottom gasification process.

- Expected equipment and materials performance is sufficiently acceptable not to preclude any of the gasification processes.

- Downstream synthesis gas processing is required for all gasification processes to produce an acceptable quality syngas for methanol synthesis. Syngas from the Koppers-Totzek, HT Winkler and Westinghouse processes require adjustment in the H_2/CO ratio whereas syngas from the Lurgi process (and to a lesser extent the Westinghouse process) require residual methane conversion to H_2 and CO.

- All effluent streams from each gasification process can be treated to meet existing environmental regulations using commercially proven technology. The magnitude of treatment facilities for the Lurgi process is more extensive than for the other processes as tars, tar oils, condensible hydrocarbons and phenolic compounds are present in the excess process water and raw synthesis gas streams.

2.2 RECOMMENDATIONS

It is recommended that all four selected gasification processes be carried forward for evaluation in Phase 1. The four processes are:

- Koppers-Totzek Entrained Bed Process
- Lurgi Dry-Ash Fixed Bed Process
- Westinghouse Fluidized Bed Process
- High Temperature Winkler Fluid-Bed Process

Specific emphasis in Phase 1 should be placed on:

- Lignite predrying requirements and the possible effect on fines generation (specifically for the Lurgi process) and on expected gasifier performance.
- Synthesis gas processing options to produce an optimum H_2/CO ratio methanol synthesis feed gas.
- Differences in effluent treatment facilities required to meet environmental regulations.

As it appears that all four gasification technologies can efficiently process Ontario lignites, differences in these key process parameters may be instrumental in selecting the gasification process(es) for testing in Phase 2.

[illegible]ABLE [illegible] SUM[illegible] GAS[illegible]ION [illegible]S CH[illegible]RIST[illegible]

	KOPPERS-TOTZEK	LURGI	WESTINGHOUSE	HIGH-TEMPERATURE WINKLER
Commerically Proven Process	Yes	Yes	No	No
Pilot Plant or PDU	No; Bench scale only 6 kg/hr; demonstration runs can be arranged at commercial facilities.	No; demonstration runs at Sasol or RUHR-100 facility in Dorsten.	Yes; 35 tonne/day at Waltz Mill, Pa.	Yes; 1 300 kg/hr at Rheinbraun plant, Frechen, West Germany. Small pilot plant at 35 kg/hr.
Experience with Lignites	Commercial Plant in Greece.	Experience with S. African brown coal and commercial test runs on North Dakota Lignite.	PDU test runs with Texas Lignite.	Yes; Rhenish Brown Coal and North Dakota Lignite.
Lignite Feedstock Size Range and Limitations	Pulverized feedstock 90% under 200 mesh; Fines can easily be handled.	Optimum size 50 mm x 5 mm Fines not recommended in fixed bed process.	Ground to 6 mm size Fines can be handled.	Fines to 12 mm size can be handled.
Gasification Thermal Efficiency/Carbon Conversion with Lignite Feedstock (Note 1)	67-75% 98%	88-95% 97%	83-86% 95%	82% 96%
Lignite Moisture Content to Gasifier	6-8% recommended	30% preferred; min. drying recommended due to fines limitations, 36% max. moisture	5-15% preferred 24% max.	8-12% preferred 18% max.
Typical Gasification Agent Consumption KG/KG Lignite Feed at Noted Moisture Content	Lignite Moisture Content: 8% Oxygen : 0.73 Steam : 0.14	Lignite Moisture Content: 30% Oxygen : 0.20 Steam : 1.10	Lignite Moisture Content: 15% Oxygen : 0.45 Steam : 0.20	Lignite Moisture Content: 8% Oxygen : 0.542 Steam : 0.167
Expected Syngas Composition using Lignite as Feedstock - % Vol. (Major Components Only)	H_2 : 29.6 CO : 55.9 CO_2 : 11.9 CH_4 : 0.1	H_2 : 38.9 CO : 16.7 CO_2 : 32.0 CH_4 : 11.0	H_2 : 21.70 CO : 35.05 CO_2 : 37.51 CH_4 : 4.53	H_2 : CO : Not Available CO_2 : CH_4 :
Material and Equipment Performance	Gasifier refractory problems due to slag attack;	No serious problems noted due to lower temperature operation and extensive commercial experience.	Material selection important due to elevated temperature (950°C) and high pressure operation (3.1 MPa)	As per Westinghouse
Downstream Processing Highlights	Extensive compression required due to atmospheric gasification; H_2/CO ratio adjustment required.	Moderate compression as gasification pressure is 3 MPa No H_2/CO ratio adjustment necessary, however excessive amount of methane produced.	Moderate compression as commercial gasifier estimated to operate at 3.2 MPa. H_2/CO ratio adjustment required.	Moderate to high compression requirements with gasifier operating at 1.1 MPa. H_2/CO ratio adjustment required.
Environmental Requirements	Air - Particulate and acid gas removal. Water - Filtration/flocculation, sour water stripping and Biox. Solids - Ash disposal.	As per Koppers-Totzek, but more extensive waste water treatment ie. oil separation and phenol removal.	As per Koppers-Totzek.	As per Koppers Totzek.

Notes 1. Comparisons should not be made between processes for thermal efficiency and carbon conversion values given, as gasification parameters and interpretation of results may vary. Also, higher values of thermal efficiency and carbon conversion for a process does not necessarily mean that the process is superior (unwanted carbon products such as CO_2 may constitute the major conversion). Overall thermal efficiency, lignite

Detailed Discussion

3.0 DETAILED DISCUSSION

3.1 GENERAL

Lignite gasification is of potential interest for the production of synthetic fuels, in particular methanol or gasoline via the Mobil Methanol-to-Gasoline (MTG) technology, as lignite reserves are substantial in Northern Ontario and conventional Canadian crude oil reserves are declining.

Lignite, when gasified in the presence of oxygen and steam at elevated temperature, reacts to form a synthesis gas containing carbon monoxide, hydrogen, carbon dioxide, methane and small amounts of hydrogen sulphide, carbonyl sulphide, ammonia, nitrogen and argon. The composition of the synthesis gas is dependent on the gasification temperature, the gasification process, the type of lignite and the impurities in both the lignite and oxygen fed to the gasifier.

Raw synthesis gas is treated to remove carbon dioxide and sulphur compounds and the resultant hydrogen and carbon monoxide-rich synthesis gas is converted to methanol via conventional technology. If synthetic gasoline is the desired end product, the methanol can be converted to gasoline via the Mobil MTG Process. This process is in an advanced stage of development and should be commercialized by 1985 in New Zealand with the startup of a 2000 m^3 per day facility utilizing natural gas as feedstock.

Lignite gasification systems can be classified according to the gasifier bed type: fixed bed, fluidized bed or entrained bed. The following are summaries of the characteristics, advantages and limitations of the three basic types:

3.1.1 FIXED BED GASIFIERS

Fixed bed gasifiers are characterized by their low operating temperature and high carbon conversion efficiency due to the countercurrent mode of operation and their low oxygen requirement for gasification. Typically represented by the Lurgi Dry Ash Process, fixed bed gasifiers are fully commercial and are well suited for highly reactive lignite when inclusion of methane in the product gas is advantageous.

Because of the low operating temperature of fixed bed gasifiers, the product gas contains tars, tar oils and condensible hydrocarbons and the process water stream (excess steam and water of reaction) contains substantial quantities of phenolic compounds, in addition to ammonia and sulphur compounds. As a result, product gas and process water streams require extensive treatment to meet product gas specifications and environmental regulations.

Fixed bed gasifiers also require sized lignite feed to prevent excessive entrainment of fines in the product gas and to prevent excessive pressure drop in the fixed bed. This is a limitation as the lignite fines (often as high as 25 percent of the sized lignite feed) must be handled separately.

3.1.2 FLUIDIZED BED GASIFIERS

Fluidized bed gasifiers provide high capacity, excellent solid-gas contact and can handle a wide range of coals and lignites, including fines. The gasifier configuration and operation is simple and tends to be stable because of the high solids inventory and the ease of introduction and withdrawal of solids. Low pressure fluidized bed gasifiers, as typified by the Winkler Fluid-Bed Gasifier, have been in commercial operation since the 1920's and have successfully processed lignites.

Recently, high pressure and temperature versions of fluidized bed gasifiers have been developed by a number of companies, including Winkler (High Temperature or HT Winkler) and Westinghouse. The pressurized version of the fluid-bed gasifiers offer all of the advantages of the low pressure version plus an increase in gas generator unit capacity and lower methane content in the product gas. The high pressure fluidized bed gasification technology is not fully commercialized at the present time but is in an advanced stage of development. A 1200 tonne per day Westinghouse gasifier is currently being engineered and constructed for SASOL in South Africa. It is scheduled to become operational in 1985.

Fluid-bed gasifiers require dry lignite for feeding and proper fluidization. Because a large volume of fluidizing media is required, fines carryover is high. In addition, carbon conversion is typically low as some of the lignite particles short-circuit the normal solid particle flow pattern in the reactor and leave with the ash.

3.1.3 ENTRAINED BED GASIFIERS

Entrained bed gasifiers have high capacity per unit volume, can efficiently process all types of lignite and coal, and produce a slagged, low carbon content ash. Because of their high operating temperature, the product gas is free of tars, hydrocarbons and phenols. The Koppers-Totzek Gasifier, typical of entrained bed gasification, is in commercial operation around the world with over 50 units in use, some of which feed lignite.

Limitations of the entrained bed gasification technology are associated with its high operating temperature, which results in high oxygen consumption, low hydrogen-to-carbon monoxide ratio in the product gas, special materials of construction and high maintenance of refractory-lined gasifiers and waste heat boilers. Also, because of the small fuel inventory, advanced control systems are required to ensure a safe reliable operation.

With the Koppers-Totzek gasifier, an additional limitation is its low operating pressure and the resultant costs associated with product gas compression to the user pressure (methanol synthesis in this case).

3.1.4 SELECTED PROCESSES

The four gasification technologies selected for evaluation in this study include two first generation processes (Koppers-Totzek Entrained Bed Gasification Process and Lurgi Dry Ash Fixed Bed Process) and two second generation fluidized bed processes (Westinghouse Gasification Process and HT Winkler Fluid-Bed Gasification Process). The first generation processes are currently in commercial use worldwide for gasifying coals and lignites and the second generation processes are in an advanced stage of development and should be commercially proven by 1995, the time frame specified for the study.

Each of these technologies is described in subsequent sections of this report.

3.2 LIGNITE EVALUATION

Tables 3.1, 3.2 and 3.3 summarize typical properties for Onakawana and Ontario lignites, Saskatchewan Coronach lignite and other lignites (or brown coals) that have been processed or tested in pilot scale or process development units by the process developers whose gasification technologies are being considered. Information on Onakawana, Ontario and Saskatchewan lignites was provided by Ontario Energy Corporation, data on North Dakota, Texas Big Brown and Rhenish (German) lignites were obtained from the literature or from process developers.

From the tables it can be seen that the data are incomplete and there are wide variations in moisture content, ash content, ash analysis and petrographic analysis between individual lignite samples from the same source. This is to be expected as chemical composition, petrographic composition and chemical reactivity typically vary considerably within a given lignite deposit, both from one location to another and across the thickness of a seam at a given location. As a result, grab samples or average samples are not necessarily representative of the entire deposit. This phenomenon is particularly valid for low rank coals and, especially, lignites.

Because of the variation in the analyses of individual lignite samples from the Onakawana and Saskatchewan Coronach deposits, it is difficult to make a valid comparison of their similarities and differences relative to efficient gasification potential. As only a partial analysis of one Ontario lignite sample was provided (no ash fusion temperature or petrographic analysis) it is not possible to make a valid comparison between this sample and the Onakawana and Saskatchewan Coronach lignites.

However, it does appear that some of the key gasification parameters are similar. Ash contents of the Onakawana and Ontario lignites appear to be similar to that of Saskatchewan Coronach lignites and ash fusion temperatures appear to be in the same range. Sodium content of the ashes also appear to be similar. Moisture content of the Onakawana and Ontario lignite appears to be 5-10 percent higher than the Coronach lignite.

Even though it is not possible to make valid comparisons between the Ontario, including Onakawana, lignites and Saskatchewan lignites, it is apparent that the Onakawana and Ontario lignites are suitable gasification process feedstocks. None of the indicated physical, chemical or petrographic properties of these lignites appear to preclude efficient gasifier operation with any of the four selected gasification processes. Initial contacts with each of the process developers have confirmed this assessment, based on their experience with similar lignites (North Dakota, Texas and German, as indicated in Table 3.1).

Testing of the Onakawana and Ontario lignites by individual process developers is required before definite conclusions can be drawn related to the suitability of each selected gasification technology to efficiently process the lignites. This is the main objective of phase 2 of the study.

Following are some general observations on the effect of lignite properties on gasification potential and thermal efficiency of the process.

3.2.1 MOISTURE CONTENT

Moisture content of the Onakawana and Ontario lignites is high, in the order of 40-55 percent. This is similar to the moisture contents of lignites processed or tested by the process developers, 30-50 percent for North Dakota lignites, 27-35 percent for Texas lignites and 35-60 percent for German lignites.

For most processes the lignite must be dried before gasification. As such, the moisture content of the run-of-mine lignite affects the overall thermal efficiency of the process in relation to the extent of predrying required. Optimum lignite moisture contents for the four selected processes are as follows:

Lurgi	30%
Koppers-Totzek	6-8%
HT Winkler	8-12%
Westinghouse	5-8%

3.2.2 ASH CONTENT

Ash content of the Onakawana and Ontario lignites, at 10-21 percent, is similar to the ash contents of North Dakota, Texas and Saskatchewan lignites but higher than Rhenish (German) brown coal.

Figure 3.1 shows that ash contents below 30 wt percent do not substantially affect the overall thermal efficiency of gasification processes. This figure was developed by Shell International for the Shell-Koppers Gasification Process but can be considered typical for the entrained bed and fluidized bed processes considered in this study. It does not apply for the Lurgi process as steam consumption increases with ash content in the Lurgi gasifier to keep the temperature in the bottom of the gasifier below the ash melting point to prevent slagging of the ash.

Lurgi gasifiers can, however, process high ash content lignites efficiently, as was demonstrated during commercial trials on North Dakota lignites at SASOL in South Africa.

3.2.3 ASH PROPERTIES

The most important ash property is ash fusion temperature as it directly affects the performance of all gasification processes. Based on the limited data available, ash fusion temperatures of the

Onakawana and Ontario lignites appear to be in the same range as those of the North Dakota, Texas and German lignites processed and tested by the process developers and similar to the ash fusion temperatures of the Saskatchewan Coronach lignite.

It should be noted that silica and alumina contents of the Onakawana, Ontario and Saskatchewan lignite ashes are higher and the calcium and magnesium contents lower than that of the North Dakota lignites tested by the process developers. This is not expected to affect gasifier performance to any significant extent, however, as ash fusion temperature appears to be the critical ash parameter.

Sodium content of Onakawana, Ontario and Saskatchewan lignite ashes, at 0.3 - 1.8 percent Na_2O, appears to be significantly lower than that of the North Dakota lignite tested by Lurgi, at 5.3 percent Na_2O. The lower sodium content should be beneficial as high sodium content reduces ash fusion temperature (in the order of 5-8°C per percent Na_2O) and increases the fouling tendency of the ash.

3.2.4 PETROGRAPHIC ANALYSIS

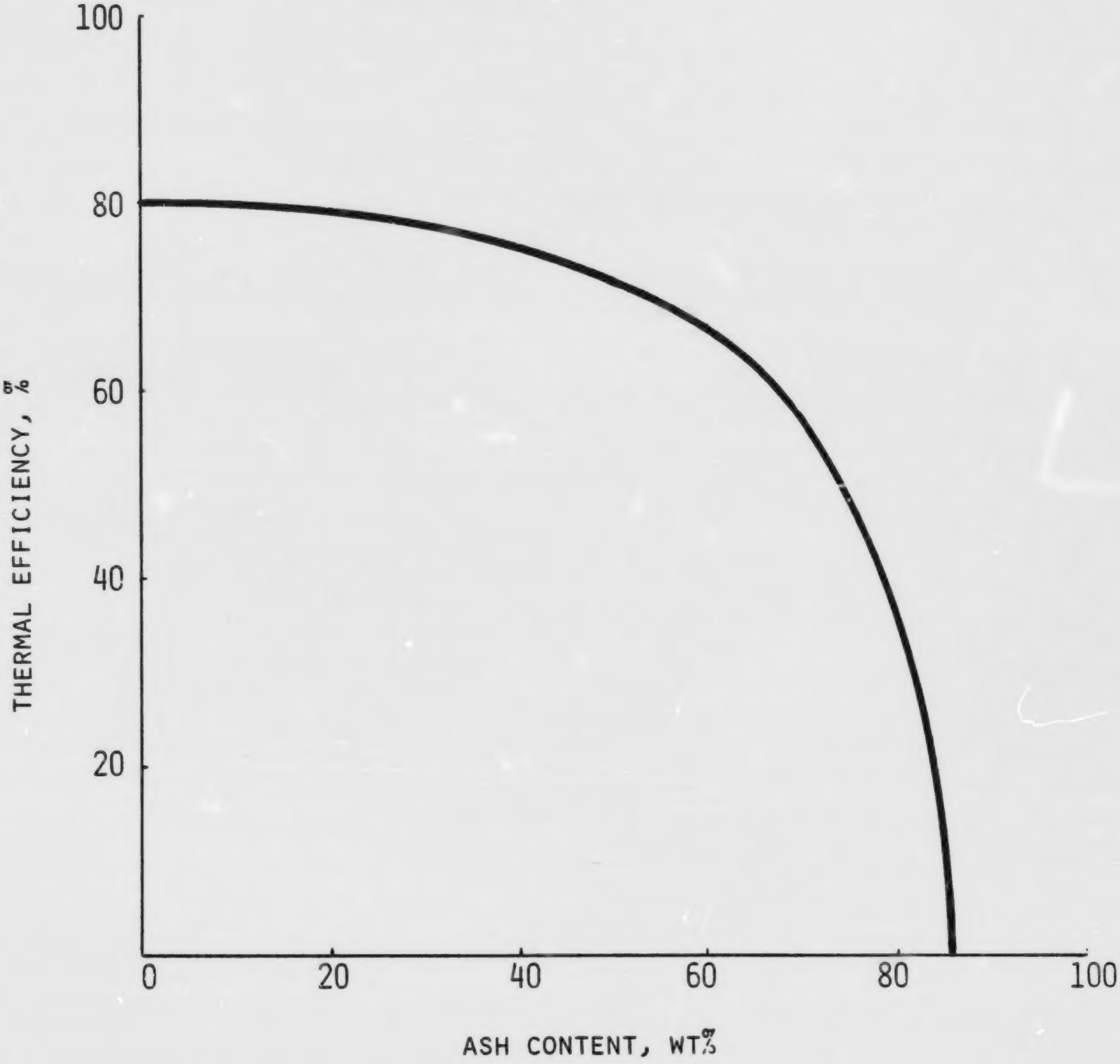
Very little is known about the behaviour of Onakawana and Saskatchewan Coronach lignites during gasification and the relationship between gasification reactivity and detailed petrographic composition. It has been observed,(1) however, that structured huminite and liptinite macerals react completely with oxygen and steam while inert macerals remain relatively unchanged and unstructured macerals react to an intermediate extent.

Based on the limited petrographic analyses and the fact that the Onakawana and Saskatchewan Coronach lignites are of different origin (Onakawana from the Cretaceous period and Coronach from the Paleocene period), differences in specific macerals were observed. The effect of these differences on gasification reactivity is unknown but is not expected to be great.

(1) Saskatchewan Power Corporation, Research and Development Centre, "Gasification and Petrographic Studies of Saskatchewan Lignites", Report on EMR Contract Serial No. 0S078-00023.

FIGURE 3.1

INFLUENCE OF ASH CONTENT ON
THERMAL EFFICIENCY FOR
OXYGEN/STEAM GASIFICATION



REF: CLEAN SYNGAS FROM COAL, M.J. VAN DER BURGT,
SHELL INTERNATIONAL, HYDROCARBON PROCESSING
VOL. 58 NO. 1, JANUARY 1978, PG 161-164.

TABLE 3.1

TYPICAL LIGNITE ANALYSES

	ONAKAWANA LIGNITE		ONTARIO LIGNITE	CORONACH LIGNITE SASKOIL #3216		NORTH DAKOTA LIGNITE	TEXAS BIG BROWN LIGNITE	RHENISH BROWN COAL
PROXIMATE ANALYSIS, WT %								
Moisture	40-55		40-45	30-40		35	16.4	8-10 (Dried)
Ash, DB	18.1		16.59	15.1		11.6	15.6	
Volatile Matter, DB	40.0		41.19	40.3		41.9	43.6	
Fixed Carbon, DB	41.9		42.22	44.6		46.5	40.8	
Heat Value, KCAL/KG, DB	5 327		5 387	5 424		5 833	5 280	5 550
ULTIMATE ANALYSIS, WT % DB								
Carbon	59.7		57.94	59.86		63.5	61.5	65.0
Hydrogen	4.1		3.84	4.33		4.2	4.6	5.0
Oxygen	19.3		19.24	20.22		19.0	16.1	24.1
Nitrogen	0.4		0.53	0.70		0.8	1.1	0.4
Sulphur	0.9		1.86	0.81		0.9	1.1	0.5
Ash	15.6		16.59	15.08		11.6	15.6	5.0
ASH FUSION TEMPERATURE, °C	Red.	Oxid.		Red.	Oxid.	Red.	Red.	
Initial Deformation	1 135	1 174		1 275	1 310	1 060	1 188	
Softening	1 191	1 235		1 310	1 340	1 088	1 238	
Fluid Temperature	1 249	1 293		1 340	1 370	1 116	1 327	
ASH ANALYSIS								
SiO_2	55.6		47.12	34.9		17.9	41.1	
Al_2O_3	11.5		23.42	24.6		11.8	11.7	
Fe_2O_3	7.2		10.01	3.90		13.2	7.6	
CaO	11.9		7.26	16.9		24.6	3.8	
MgO	3.2		1.85	4.51		6.7	2.4	
TiO_2	1.0		1.19	1.27		1.82		
Na_2O	1.04		0.26	0.61		5.3	0.02	
K_2O	1.42		0.60	0.97		0.33	0.44	
P_2O_5	-		0.24	0.13		0.65		
BaO	0.26		0.08	-		-		
SrO	0.13		0.17	-		-		
SO_3	8.40		3.98	10.7		17.7	24.9	
LGI (Misc)	0.41		2.6	0.11		-		
PETROGRAPHIC ANALYSIS, VOL %								
Huminite - Structured				55.0				
- Groundmass				16.4	73.4			
- Unstructured				2.0				
Liptinite				2.0				
Iner[illegible]				[illegible]2.6				

TABLE 3.2

ONAKAWANA AND ONTARIO LIGNITE ANALYSES

	ONAKAWANA LTH-01 PRELIMINARY		ONAKAWANA LTH-03 SAMPLE #1	ONAKAWANA LTH-03 SAMPLE #2	ONAKAWANA SNC/ER ON-4		ONAKAWANA SNC/ON		ONTARIO
PROXIMATE ANALYSIS, WT %									
Moisture	49.9		-	-					40-45
Ash, DB	18.1		16.79	20.62					16.59
Volatile Matter, DB	40.0		41.87	39.12					41.19
Fixed Carbon, DB	41.9		41.34	40.06					42.22
Heat Value, KCAL/KG, DB	5 327		5 506	5 168					5 387
ULTIMATE ANALYSIS, WT % DB									
Carbon	59.7		60.69	57.92					57.94
Hydrogen	4.1		3.94	3.68					3.84
Oxygen	19.3		16.64	15.29					19.24
Nitrogen	0.4		0.67	0.74					0.53
Sulphur	0.9		1.27	1.55					1.66
Ash	15.6		16.79	20.82					16.59
ASH FUSION TEMPERATURE, °C	Red.	Oxid.							
Initial Deformation	1 135	1 174							
Softening	1 191	1 235							
Fluid Temperature	1 249	1 293							
ASH ANALYSIS									
SiO_2	55.6		39.97	39.11					47.12
Al_2O_3	11.5		12.01	11.45					23.42
Fe_2O_3	7.2		9.63	10.77					10.01
CaO	11.9		15.56	14.24					7.26
MgO	3.2		4.29	3.68					1.85
TiO_2	1.0		0.89	0.85					1.19
Na_2O	1.01		1.34	1.09					0.26
K_2O	1.42		1.01	0.77					0.60
P_2O_5	-		0.13	0.16					0.24
BaO	0.15		-	-					0.08
SrO	0.13		0.13	0.14					0.17
SO_3	8.40		14.93	16.20					3.98
LOI (Misc)	0.41		1.01	0.80					2.6
PETROGRAPHIC ANALYSIS, VOL %									
Huminite - Structured					74.2		52.2		
- Groundmass					4.2	94.6	16.0	73.0	
- Unstructured					16.2		4.8		
Liptinite					1.4		2.4		
Inertinite					1.4		20.4		

3.3 GASIFIER TECHNOLOGIES REVIEW

The intent of Phase 0 of this study is to select the gasification technologies that are suitable to process Ontario Lignite including Onakawana Lignite. These processes will be subject to further study during Phase 1. Accordingly, the gasifications technologies are reviewed in this section as follows:

- General documentary of technology describing development, pilot plants, process demonstration units and commercial units in operation and experience with lignites.
- Brief process description.
- Feedstock requirements, particularly with respect to lignite.
- Process performance outlining carbon conversion, thermal efficiency, typical raw gas composition (with lignite if information is available).
- Operating performance outlining plant reliability, flexibility, equipment and material performance.
- Process developers initial comments as received in March 1983 with respect to suitability of process to handle Ontario and Onakawana lignites and recommended tests.
- Downstream process flexibility where each process is evaluated and the routes recommended to achieve methanol as the primary end-product.
- Environmental considerations where the waste water, gaseous effluent and solid waste applicable to each process are discussed in general terms related to the Federal and Provincial requirements.

3.3.1 KOPPERS-TOTZEK GASIFICATION PROCESS

3.3.1.1 Introduction

The Koppers-Totzek process is a fully entrained suspension gasification process which operates at atmospheric pressure. The gasification process can produce a synthesis gas from a wide variety of fuels including bituminous and sub bituminous coals, lignite, peat, heavy petroleum residue and petroleum coke.

The process has been commercially proven since 1952, at which time three gasifiers were installed at Oulu, Finland. Since then an additional 18 plants (50 gasifiers) have been built or are under construction around the world. The gasifiers are now licensed by Krupp-Koppers, a subsidiary of Krupp, Essen, West Germany. The majority of the commercial plants produce synthesis gas for ammonia manufacture. Three plants, in France, Zambia and South Africa produce methanol as well as ammonia.

A Koppers-Totzek process using lignite as a feedstock is in operation in Ptolemais, Greece producing synthesis gas for ammonia synthesis. A high-ash lignite (30-35%) dried to 8% moisture content is processed in six gasifiers.

Features of the Koppers-Totzek process include:

- A wide variety of feedstocks can be handled and the process is flexible with respect to varying coal properties.
- No byproducts such as tars, phenols, or condensible hydrocarbons are produced.
- Equipment is simple and operating capacity is flexible; full capacity can be attained within 30 to 60 minutes of startup.

Some of the drawbacks include:

- Operation at atmospheric pressure and low capacity per gasifier; also there is a high cost of compression for subsequent processing.
- Behaviour of ash at high temperatures in refractory lined gasifier may cause severe problems.

3.3.1.2 Process Description

A block diagram describing the production of synthesis gas is described in Figure 3.2.

Pulverized coal, 90% under 200 mesh, is screw fed from hoppers into twin burners. The coal ranges from 1-8% moisture content, depending on the type of feedstock. The gasifier is shaped like two spheroidal cones and centrally welded together. The gasifier is refractory lined with a water jacket in which process steam is generated and added to the burner nozzles in quantities sufficient to maintain the flame temperature at about 1900°C. The gasification is performed virtually immediately (less than 1 second) resulting in a temperature of the produced gas of 1400-1600°C, so that all the hydrocarbons contained in the coal are converted into CO, CO_2 and H_2. About 50% of the ash leaves as slag, to be quenched and granulated in water at the bottom of the gasifier, while the other 50% leaves with the product gas. The product gas exits the gasifier and passes through a waste heat boiler generating high pressure steam and then passes to a quench unit to remove the fly ash.

The gas exits the spray washer at 35°C and is cleaned further in two disintegrators arranged in series. Exiting the disintegrators, the gas passes through a mist eliminator to remove entrained moisture.

The produced gas is at atmospheric pressure and must be compressed for further treatment. Electrostatic precipitators may be required to treat the gas further to protect compression equipment.

3.3.1.3 Feedstock

The use of both high temperature and pulverized coal in the Koppers-Totzek gasifier results in the ability to handle a variety of feedstocks and the type of coal has only very little influence on the gas composition. Lignite, when used as feedstock, is dried to about 8% water content and pulverized to a size 90% under 200 mesh.

3.3.1.4 Process Performance

The Koppers-Totzek gasifier has a thermal efficiency of between 67 and 75%. Carbon conversion is approximately 98% for lignite. A typical gas composition using lignite as a feedstock is as follows:

	% Vol.
H_2	29.6
CO	55.9
CO_2	11.9
N_2+Ar	2.0
CH_4	0.1
H_2S/COS	0.5

3.3.1.5 Operating Performance

Operating performance data, based on experience gained at AECI Ltd., Modderfontein, South Africa, are discussed here. In this plant, six double headed Koppers-Totzek gasifiers produce synthesis gas for ammonia manufacture.

The major causes of breakdown during the first year of operation were mechanical in nature, particularly in the waste heat boilers. Tube and grate failures were persistant problems, which contributed significantly to plant shutdowns or rate reductions. These in turn generated mechanical problems throughout the plant due to intermittent operation, start-ups and emergency shutdowns.

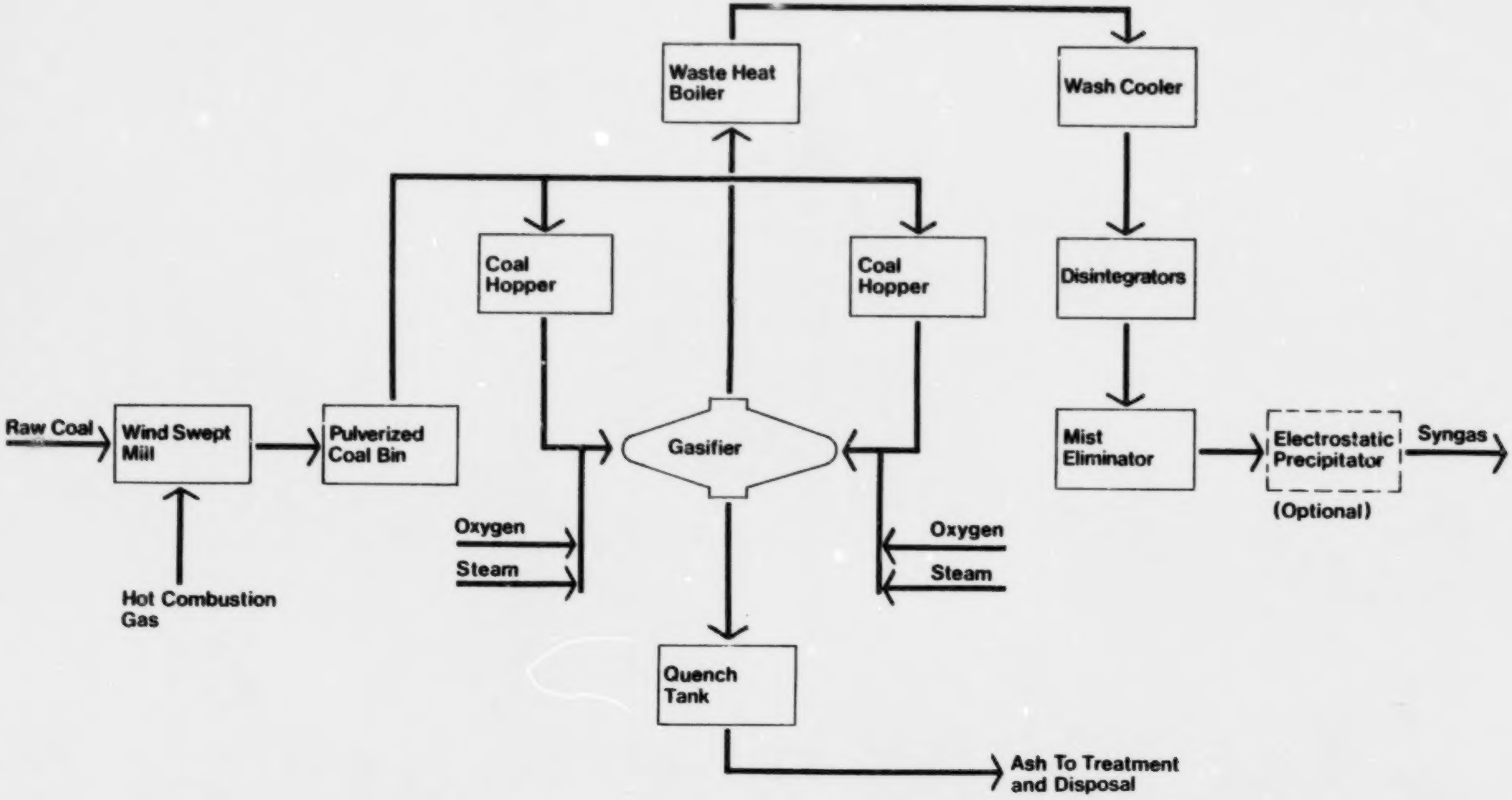
This phase of mechanical problems was eventually overcome. During periods of continuous operation a more detailed assessment could be made.

The gasifiers were capable of producing the rated output of gas with a composition very close to predicted. However, four main difficulties were identified as follows:

1) Loss of the gasifier refractory lining.
2) Low carbon conversion efficiency.
3) Excessive generation of carbonaceous fly ash.
4) Fouling of raw gas compressors.

Loss of the lining was caused by slag attack due to too high a temperature operation. The gasifier temperature was reduced which corrected the problem but reduced carbon conversion efficiency. Modifications were also made to the refractory lining and the cooling jacket which have operated successfully since the changes.

The low carbon conversion efficiency was caused by the feedstock being unsuitable due to very low reactivity and poor mineralogical characteristics. In addition the poor mineralogical characteristics led to very little slag formation during gasification. Consequently, the bulk of the coal left as fly ash.


Raw Coal
Wind Swept Mill
Hot Combustion Gas
Pulverized Coal Bin
Coal Hopper
Oxygen
Steam
Waste Heat Boiler
Gasifier
Quench Tank
Coal Hopper
Oxygen
Steam
Ash To Treatment and Disposal
Wash Cooler
Disintegrators
Mist Eliminator
Electrostatic Precipitator
(Optional)
Syngas

Koppers - Totzek Process
Figure 3.2

3.3.2 LURGI PROCESS

3.3.2.1 Introduction

The Lurgi first-generation process is a dry-bottom fixed bed system developed in the late 1920's for German brown coal. Lurgi process gasification uses a mixture of steam and oxygen (and/or air) as gasification agent operating at pressures up to 3 100 kPa (450 psi). All grades of coal can be processed, but the process is more suitable for low rank coals such as lignite, sub bituminous coal and brown coal.

The first commercial plant was built in 1936, and the process has since been continuously developed and refined as gasification technology advanced. A total of 16 plants exist comprising over 100 gasifiers in use or on order. The most significant commercial usage of Lurgi gasifiers has been in South Africa where SASOL operate two coal to gasoline plants via the Lurgi gasification process and are presently commissioning a third plant. SASOL I contains 16 operating gasifiers with one of 5 m diameter, the largest Lurgi gasifier yet constructed. SASOL II contains 36 operating gasifiers with SASOL III projected to operate with a further 36 gasifiers.

In the U.S.A. a major commercial venture is in the development stage for the construction of a 22 000 t/day lignite to SNG plant in North Dakota. The plant will use the Lurgi gasification process at a pressure of 3 MPa (430 psig) to produce a raw gas which will be further processed to provide a pipeline quality SNG with higher heating value of 37 MJ/Nm^3 (977 BTU/SCF). This venture is being managed by a consortium, the Great Plains Gasification Associates (GPGA). North Dakota Beulah-Hazen Lignite was extensively tested by Lurgi in the laboratory and encouraging results led to a full-scale test in the SASOL plant in South Africa using 12 000 tonne of the lignite, shipped from North Dakota. The Great Plains process and lignite testing is discussed in more detail in subsequent sections.

3.3.2.2 Process Description

A simplified block diagram of the proposed Great Plains Gasification process is shown in Figure 3.3. The primary product for this process is SNG. However, if methanol was the primary product, shift conversion would not be required.

Lignite, (22 000 t/day) is crushed in primary and secondary crushers to a gasifier feedstock of size 50 mm x 5 mm. Fines below 5 mm size are rejected at a rate of 8 000 t/day and fed to a nearby power station. The gasifier building contains 14 Lurgi Mark IV gasifiers. Twelve are sufficient for design capacity with two spares allowing a continuous maintenance program without reducing output. A typical gasifier is shown in Figure 3.4. The gasifier is a pressure vessel with water jacket with the lignite moving countercurrent to the gasification agent. Lignite is charged through the top by means of lock hoppers and a coal distributor. Oxygen (from the air separation plant) and steam are injected through a rotary grate as gasification agents and the steam also serves to cool the ash to maintain it below its melting point. The ash is removed through the rotary grate and discharged via lock hoppers. The gasifiers operate at 3 MPa (430 psig).

The raw gas exiting the gasifiers is quenched and cooled to condense the tar, oil, phenols, ammonia and water from the gas stream. These components comprise the gas liquor which is treated separately. About 30% of the gas is passed through a shift converter to produce the correct stoichometric H_2/CO ratio for methanation. The streams then combine to be passed through a Rectisol Unit (a low temperature methanol wash) to remove all naphtha, essentially all sulfur and over 97% of the CO_2.

The clean synthesis gas is diluted with recycled methanated gas and passed through a methanation unit to produce SNG of HHV 37 MJ/Nm^3 (977 BTU/SCF).

The Rectisol off gases are scrubbed in a Stretford unit producing elemental sulfur. The gas liquor is processed to remove anhydrous ammonia for sales. The tar, oil, phenols and naphtha also produced are used to fuel the plant boilers and superheaters.

3.3.2.3 Feedstock

The Lurgi gasification process is more suitable for low rank coals such as lignite, as these coals are sufficiently reactive at low temperature. The process is not suitable for high rank coal such as bituminous because the higher temperature required for gasification demands the use of excess steam to prevent sintering. Furthermore, the direct feeding of reactive bituminous coal has not been commercially demonstrated in the Lurgi unit. Also, the moving-bed unit requires a coarse-sized feed greater than 5 mm. Particles finer than 5 mm are not usable in the Lurgi gasifier and must be handled separately. In the Great Plains plant about 35% of the lignite mined does not meet the size criteria and is consumed by a neighbouring power plant.

3.3.2.4 Process Performance

A typical Syngas composition before treating using lignite as feedstock is as follows:

	% Vol.
H_2	38.9
CO	16.7
CO_2	32.0
CH_4	11.0
Cn Hm	1.0
N_2+Ar	0.2
H_2S/COS	0.2

Overall plant thermal efficiency for the Lurgi process has been reported as follows:

SNG	(Based on HHV)	68.0%
SNG	(Based on LHV)	63.7%
Methanol	(Based on LHV)	51%
Methanol + SNG	(Based on LHV)	60%

In order to discuss in more detail the process performance, results of tests performed by American Natural Gas Service Company (subsequently Great Plains Gasification Associates) on the North Dakota Lignite are presented below.

Approximately 12 000 tonnes of lignite were shipped to SASOL'S Sasolburg Plant in South Africa for full scale plant trials in September 1974. The test program commenced September 5, 1974 and was completed September 17, 1974. The tests demonstrated that the Lurgi process could gasify North Dakota Lignite efficiently and obtain a gas of composition suitable for subsequent processing to the end-products desired by A.N.G.S.

The following are observations made as a result of the test:

- The moisture content of the lignite as received was about 36%. Even at this moisture content the gas outlet temperature was still above the dew point by about 83°C. This meant that pre-drying was not necessary for North Dakota lignite.
- A Mark IV gasification capacity of 57 000 Nm^3/hr could be attained and continuously produced using North Dakota lignite.
- Oxygen and steam consumption per Nm^3 of crude gas were lower for lignite than other U.S. coals tested on the Lurgi gasifier with a significant yield of methane for each type of coal.

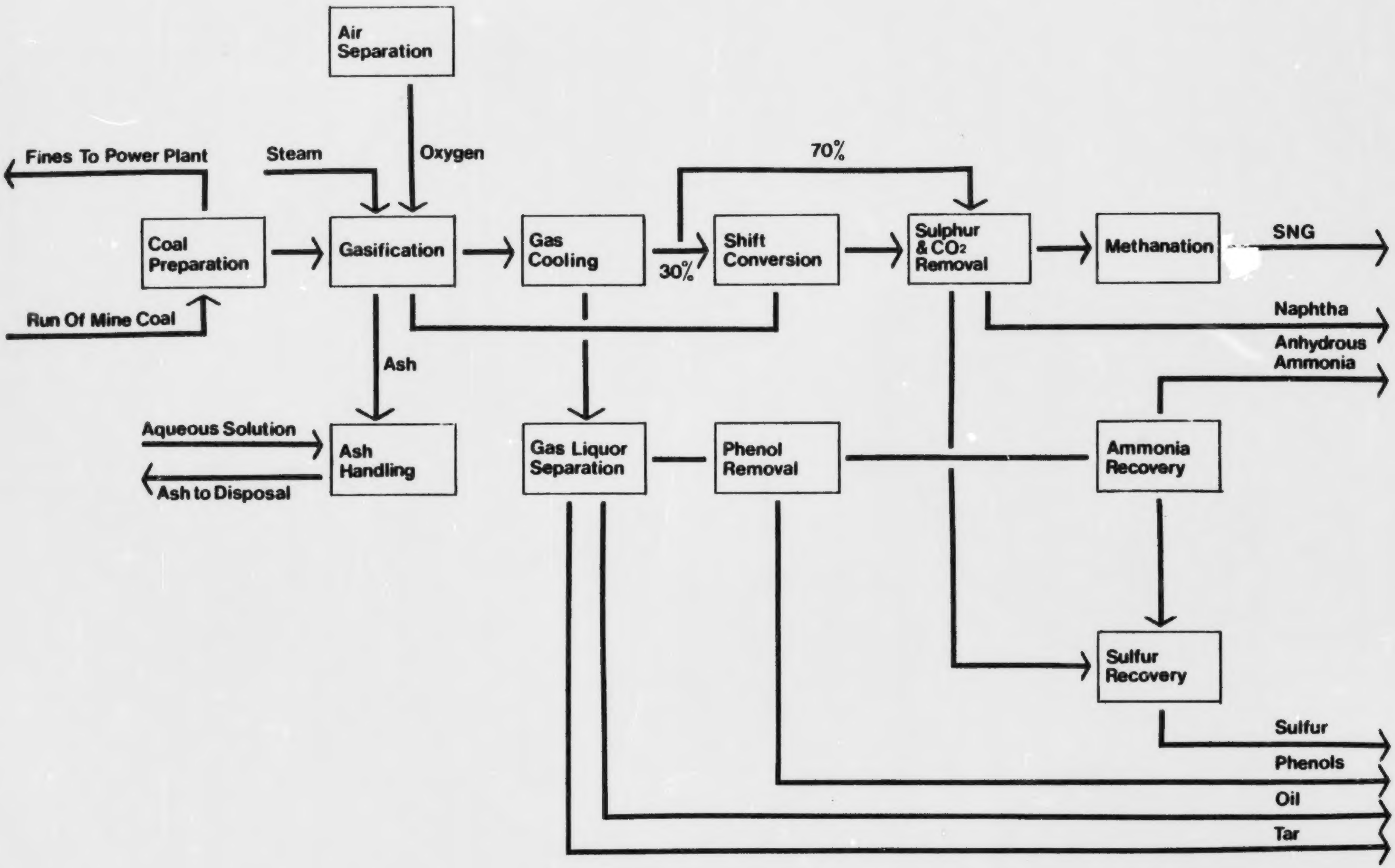
  Typical gasification agent consumptions and methane yields are as follows with comparison provided using other coals:

	Nm^3 O$_2$	m^3 HP-steam	Nm^3 CH$_4$
	per Nm^3 crude gas		per tonne d.a.f. coal
North Dakota Lignite	0.123	1.05	199
Rosebud High Volatility Coal	0.126	1.026	201
Illinois No. 5 and No. 6 graded	0.181	1.337	180
Pittsburgh No. 8 graded	0.197	1.493	186

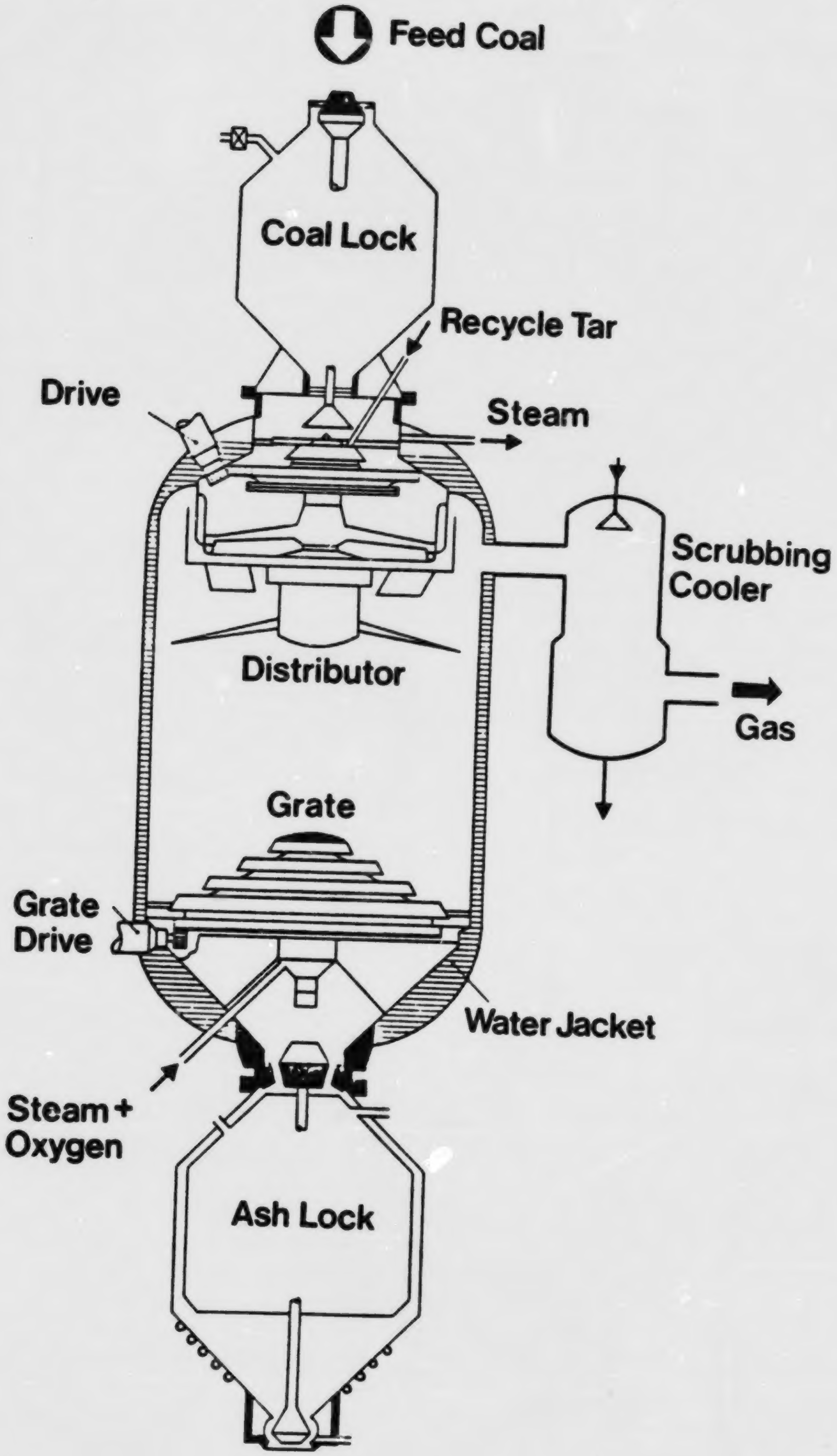
3.3.2.5 Operating Performance

The Lurgi pressure gasifier has been continuously developed since 1936. Over the years, successive improvements in mechanical features and operational techniques have eliminated virtually all major problems that can occur. When testing the North Dakota lignite, problems were encountered with the ash having a tendency to agglomerate, but this was overcome by making modifications to the grate.

The Lurgi process has been subject to the traditional coal handling problems common to all gasification processes such as coal lock-hopper and ash removal locks but these are now under control.


Air Separation
Fines To Power Plant
Steam
Oxygen
70%
Coal Preparation
Gasification
Gas Cooling
30%
Shift Conversion
Sulphur & CO2 Removal
Methanation
SNG
Run Of Mine Coal
Naphtha
Ash
Anhydrous Ammonia
Aqueous Solution
Ash Handling
Ash to Disposal
Gas Liquor Separation
Phenol Removal
Ammonia Recovery
Sulfur Recovery
Sulfur
Phenols
Oil
Tar

Lurgi Gasification Process
Great Plains Gasification Associates Process
Figure 3.3


Feed Coal
Coal Lock
Recycle Tar
Drive
Steam
Scrubbing Cooler
Distributor
Gas
Grate
Grate Drive
Water Jacket
Steam + Oxygen
Ash Lock

Lurgi Pressure Gasifier
Figure 3.4

3.3.3 THE WESTINGHOUSE PROCESS

3.3.3.1 Introduction

Westinghouse Electric Corporation has been developing since the early 1970's, a pressurized fluidized bed gasification technology initially with the support of the U.S. Department of Energy and more recently the Gas Research Institute. Westinghouse has operated a nominal 35 tonne/day process demonstration unit (PDU) at Waltz Mill, near Pittsburgh, Pennsylvania since 1975. The PDU has been operated for more than 8 000 hrs since 1975, testing a wide variety of coals and chars ranging from 15 t/day with air to 35 t/day with oxygen. The system has processed run-of-mine coal, reactive U.S. Western lignites, sub bituminous coals, caking U.S. eastern, mid-western and high ash coals. The process produces virtually no tars in the product gas, and high carbon conversion has been reported to be achieved with low use of oxidant and steam. Normal pilot plant operation is at 1 687 kPa (230 psig). The overall objective of the development program is to demonstrate the viability of the Westinghouse process for the production of low and medium BTU gas for electrical power generation, syngas, chemical feedstocks and industrial fuels.

Westinghouse is presently pursuing projects which will lead to commercial demonstration of the fluidized bed gasification technology. The pilot plant is being continuously developed to provide a data base sufficient for the design of commercial scale hardware. Heat recovery units, heat exchangers, cyclones have been added to test various operating configurations.

A cold flow model test facility, 3 m in diameter, has been designed and constructed to study the effect of scale (size) on the important solids flow behavior and gas-solids contacting in the gasifier. The model provides a tool for corroborating models, developed on the process development unit and in the laboratory, for defining commercial scale component designs.

The most recent development in the gasification process scale-up program is that SASOL Ltd., South Africa and Westinghouse have jointly undertaken to construct a full-scale, 1 200 t/day coal gasification demonstration plant at the SASOL II coal to synthetic fuels facility in Secunda, South Africa. This plant will consist of a coal preparation and feed system, a gasifier, an ash removal system, a fines collection and recycle system and heat recovery and gas quench systems. Engineering commenced in 1982 with mechanical completion scheduled for late 1984.

3.3.3.2 Process Description

A block flowsheet of the Westinghouse process describing the Waltz Mill PDU is shown in Figure 3.5.

Coal is received in lump form and prepared in an impact mill crusher dryer providing a feedstock of 5 mm size with no more than 10% smaller than 100 mesh, and dried to a free-flowing condition. Coal feeding is performed by pressurized lock-hoppers and pneumatic transport of feedstock using recycled product gas as the transport medium.

The gasifier consists of four functional zones: (Refer Fig. 3.6)

(1) An expanded freeboard section to separate char fines from exiting raw gas.

(2) A gasifier bed to permit char particles to react with steam.

(3) A combustor to devolatilize the coal, combust carbon for the heat source, and produce ash agglomerates.

(4) A char-ash separator to separate the ash agglomerates from char, to cool the ash with recycle gas and to discharge ash into the lock hoppers.

Typical operating ranges for the unit are as follows:

Freeboard Temperature	900 - 1 000°C
Pressure:	1 000 - 1 687 kPa (130 - 230 psig)
Coal Flow Rates:	450 to 730 kg/hr for bituminous coals 730 to 1 600 kg/hr for sub-bituminous coals and lignites

Fines-laden raw gas from the gasifier flows through two cyclones to remove entrained particulates. The solids are recycled to the gasifier using recycle gas as the transport medium. The raw gas is cooled in a heat recovery system generating superheated steam, and then quenched in the Gas Quench System to remove any remaining particulates entrained in the gas. A slurry is continuously drained from the tower bottom to the pressure filtration system. The product gas leaving the gas scrubbing tower is split into two streams, one to downstream processing while the other is used as recycle gas for pneumatic transport.

3.3.3.3 Feedstock

Westinghouse has tested sub-bituminous and bituminous coals, chars and lignites in their pilot plant. The lignite most extensively tested was Texas lignite during 1980-1981. The total duration of the tests was 257 hours of which 113 hours were conducted on a test run in August 1981 on a lignite from Big Brown Mine, Texas.

Table 3.4 gives the feedstock properties of the Big Brown Lignite with typical Onakawana analyses included for comparison:

TABLE 3.4

	Big Brown	Big Brown Moisture Free Basis	Onakawana Average of Samples #1 and #2
Proximate Analysis (wt%)			
Volatile Matter	35.16	41.14	40.5
Fixed Carbon	32.18	37.66	40.7
Moisture	14.54		
Ash	18.12	21.20	18.8
Ultimate Analysis (wt%)			
Carbon	48.59		59.30
Hydrogen	5.33		3.81
Oxygen	25.60		15.96
Nitrogen	1.14		0.71
Sulfur	1.22		1.41
Ash	18.12		18.81
Elemental Analysis (wt% of ash)			
SiO_2	51.25		39.54
Al_2O_3	11.06		11.73
CaO	6.37		14.90
MgO	1.80		3.99
Na_2O	0.01		1.22
K_2O	0.70		1.22
Fe_2O_3	7.40		10.20
Initial Deformation Temp °C (Reducing athmosphere)	1 187		
Softening Temperature (Reducing ath) °C	1 243		
Fluid Temperature (Reducing ath) °C	1 415		

Although in all tests the feedstock was screened through a 6 mm screen, a notable feature of the lignites tested was that storage greater than two months caused decrepitation causing additional fines to be generated, lowering the average particle size from 1 100 µm to 850 µm. Moisture content of feedstock lignite can range from 5-15% but 24% moisture content feed has been successfully processed.

3.3.3.4 Process Performance

Tests were conducted by Westinghouse on Texas lignite at various process conditions to obtain reliable data for use in scale-up design of the gasifier. Major data and observations include:

- Texas lignite feestocks have been successfully processed in the oxygen blown PDU gasifier. Typical gasifier conditions were: Pressure 1 000 - 1 500 kPa; Temperature 815-890°C, Steam/lignite ratio: 0.2 to 0.5, Lignite feed rates: 500-1 400 kg/hr; Carbon conversion up to 95% allowing for optimum fines recycle.

- A medium BTU gas was produced. A typical gas composition, dry (%) is:

H_2	21.70
CO	35.05
CO_2	37.51
CH_4	4.53
H_2S	0.28

- In order to achieve greater than 90% carbon conversion it is necessary to maintain the ash content in the discharge stream at 85% and restrict the fines losses from the cyclones to less than 5% of the lignite feed.

3.3.3.5 Operating Performance

Westinghouse has an established set of procedures and checks to be followed during a test program in order that the PDU can be operated in a safe and efficient manner and to develop process data.

For a typical experimental test such as that for Texas lignite the following sequence is followed:

1) Pretest Preparation

Pretest preparation consists of preparation of the test plan and procedure. A failure action plan is prepared which defines the step to be followed in the event of a failure. During this phase, feedstock inventory is checked and lab analyses performed to ensure size cut and chemical analysis meet specification. Delivery of utility gases required for the test, such as carbon dioxide, propane and oxygen is scheduled. Pretest briefings are held, and a pretest check list procedure to check out equipment and instrumentation is implemented.

2) Pressurization and Leak Test

The gasifier is pressurized to 1 500 kPa in 350 kPa increments to detect leaks, and following any repair the system is blocked in. If the leak rate is less than 200 kPa per hour the test is continued.

3) Refractory Hot Air Dryout

The temperature is increased by 10°C per hour until 260°C is achieved in the freeboard section of the gasifier. This temperature is held for approximately 12 hours.

4) Solids-Feed Initiation

Coke breeze (used as start-up feedstock) is transferred to the lock hoppers, and recycle gas flows started.

5) Ignition and Temperature Ramp

Coke breeze feedstock is introduced to the gasifier and the temperature ramped to combustion temperature and finally to 870°C where a hold is initiated for 4 hours. After the 4 hours, lignite feed is initiated.

6) Set Point Conditions

These conditions are declared once temperatures and flow conditions specified in the test plan are achieved. Log books are maintained to record process conditions, anomalies and daily activities. Meetings are held daily to review test status.

7) Shutdown

After the goals of the test are achieved, shutdown is initiated, and equipment depressured and cooled.

8) Post-Test Inspection

Upon conclusion of a test, disassembly procedures are initiated photographs taken and inspection reports are written.

9) Data Review and Test Report

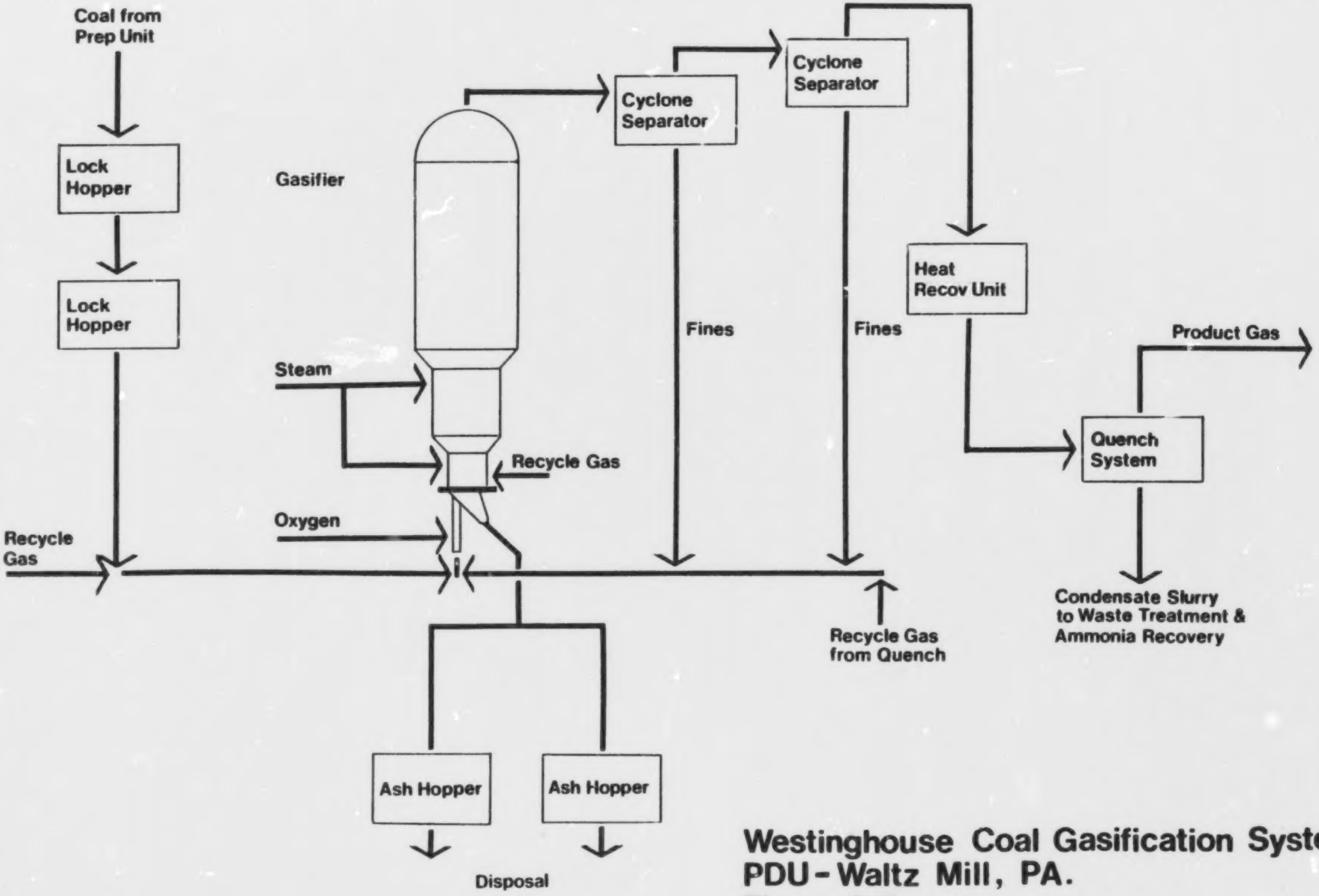
After the test the data are reviewed in detail and steady state test points are declared. A steady state is defined as that period during the set point in which all flow and pressure parameters fall within ±5% of average value and all temperatures hold to ±2.5% of average value. A heat and material balance is performed on the as-measured data, and compiled into a final report.

In the actual lignite test program the sequence described above was followed and the steady state conditions achieved in order to obtain meaningful test data.

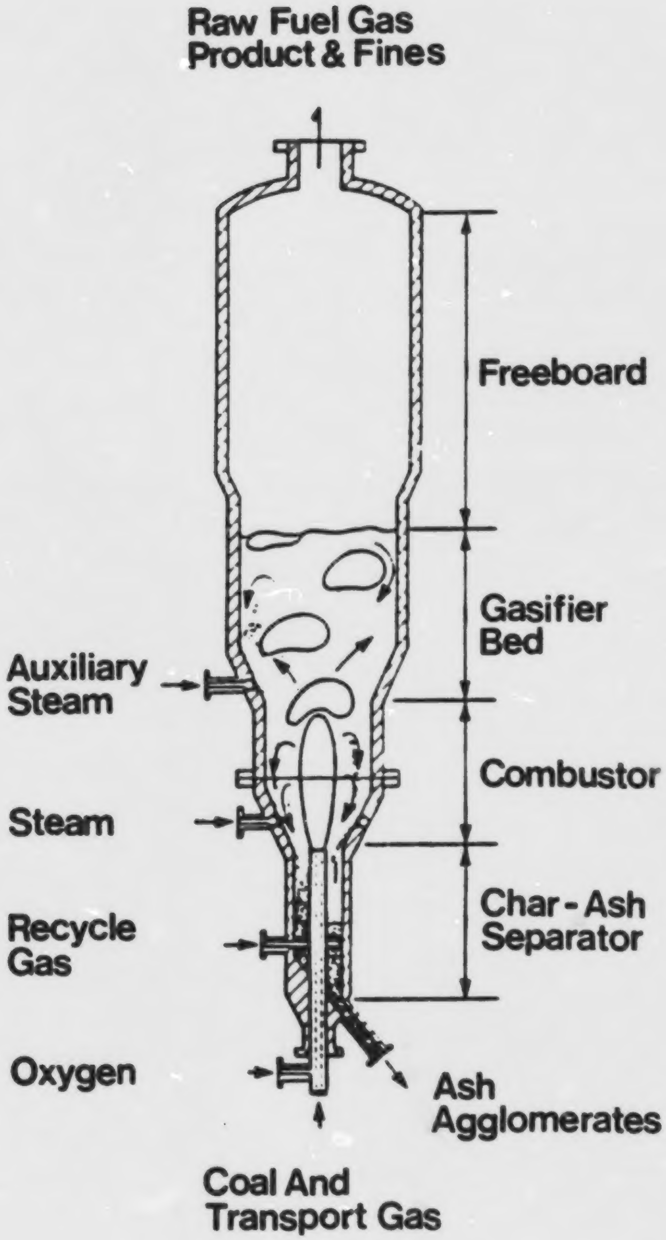
During the some 8 000 hours of operation, Westinghouse has gained experience in the operation of a gasification unit at varying levels of operation using different feedstocks resulting in continuous modifications and improvement to operating hardware. Some of the problems encountered and modifications made are as follows:

- Lock hopper load cell malfunction due to imbalanced load, improved by redesign of load cell.
- Solids flow rate by starwheel feeders, improved by purging with carbon dioxide.
- Pneumatic conveying lines blockage, improved by installing back-up systems.
- Pneumatic line erosion, improved by installing a unique erosion-resistant feed line elbow.
- Thermocouple and thermowell assemblies failing due to hydrogen sulphide attack. This situation was remedied by changing materials of construction.
- Recycle gas and process air compressors frequently needed maintenance and repair and modifications were made to improve the efficiency and dependability of these units.
- Modifications and improvements have also been made to the product quench and scrubbing options and the gasifier and cyclones refractory lining.

The PDU gasifier has shown its operational flexibility and reliability to routinely withstand failures such as loss of steam and/or oxygen, loss of coal feed, total utilities outage and loss of power. During the Texas lignite test run a total power outage due to lightning was experienced for 15 minutes, with the major effect of defluidizing the bed and reduction of temperature from 815°C to 650°C. When power was restored, temperature was ramped up, feeds resumed and gasifier operation continued smoothly for 47 hours without any adverse impact because of the slumped bed condition.


Coal from Prep Unit
Lock Hopper
Lock Hopper
Recycle Gas
Gasifier
Steam
Recycle Gas
Oxygen
Cyclone Separator
Cyclone Separator
Fines
Fines
Heat Recov Unit
Quench System
Product Gas
Condensate Slurry to Waste Treatment & Ammonia Recovery
Recycle Gas from Quench
Ash Hopper
Ash Hopper
Disposal

Westinghouse Coal Gasification System
PDU - Waltz Mill, PA.
Figure 3.5


Raw Fuel Gas
Product & Fines
Freeboard
Gasifier
Bed
Auxiliary
Steam
Combustor
Steam
Char - Ash
Separator
Recycle
Gas
Oxygen
Ash
Agglomerates
Coal And
Transport Gas

Westinghouse Gasifier
Figure 3.6

3.3.4 HIGH-TEMPERATURE WINKLER PROCESS

3.3.4.1 Introduction

The original Winkler gasification process was developed in Germany in 1922 by F. Winkler. The fluidized bed reactor for gasification showed marked advantages compared with existing processes already known. The Winkler unit was capable of a substantial increase in capacity and was able to process a wide variety of coals. The first commercial scale plant was built at Leuna, Germany, in 1926. Since then Winkler gasifiers have been built and operated in Spain, Yugoslavia, Japan, Turkey, India and West Germany. Experience shown in gasification plants has demonstrated the following capabilities of the Winkler process.

- Simple feedstock preparation, as a wide grain size range can be used.
- Low consumption of gasification agents i.e. oxygen and steam.
- Satisfactory partial load behavior over a wide performance range.
- Simple start-up and shutdown procedures.

From 1956 to 1964, two Winkler fluidized bed gasifiers with a capacity of 17 000 Nm^3/h (about 600 000 SCFH) each were operated by the Union Rheinische Braunkohlen Kraftsoff AG, a subsidiary of the Rheinische Braunkohlenwerke AG [Rheinbraun]. In these plants Rhenish brown coal, a lignite, was gasified.

Due to the changing energy supply picture, Rheinbraun further developed the fluidized bed technology to increase the advantages of the conventional Winkler process. The main areas of further development carried out jointly by Rheinbraun and Uhde are:

- Gasification under pressure aimed at increasing specific output and decreasing the compression energy demand. Conventional Winkler gasifiers operate at slightly above atmospheric pressure while the developed version operates at 1 100 kPa (160 psi) max.

- Gasification at elevated temperatures for improving the gas quality and increasing the conversion rate, thus the High-Temperature Winkler process. Conventional Winkler units operate at below the melting point of the ash (800°C for lignite, 1100°C for bituminous), so is most suited to highly reactive coals, i.e. lignite or reactive hard coals. The High-Temperature unit when processing lignite, has been operated as high as 1100°C. The melting point of the ash has been controlled by the addition of limestone to the feedstock.

- Increasing the carbon conversion rate by partially recycling the flue dust to the gasifier.

Therefore, it is not only a temperature increase that has characterized the High-Temperature Winkler process but a pressure and recycle rate increase as well.

Rheinbraun performed bench scale tests at Aachen University and then decided to build a pilot plant which was commissioned in mid-1978. This facility is located at the site of the Wachtberg plant in Frechen, near Cologne, West Germany. Experience at this plant has shown the process to be highly efficient, flexible and easy to operate. The High-Temperature Winkler process can be used in the production of high or low BTU gas, reducing gas or synthesis gas. The major design data of this plant are as follows:

Coal Throughput	:	1 300 kg/hr max.
Gasification Agents	:	Oxygen or air, and steam
Gas Output	:	Up to 2 200 Nm^3/hr
Pressure	:	1 000 kPa
Temperature	:	Up to 1100°C

The promising results achieved from the some 8 000 hours of operation of the pilot plant prompted Rheinbraun's decision to install the first commerical size synthesis gas plant employing the High-Temperature Winkler process. The plant will be located at the site of Rheinbraun's lignite briquetting plant at Berranrath with the primary product being synthesis gas. The gas will be piped to the Union Kraftstoff plant, Wesseling for the production of methanol. Engineering work started in 1981, site preparation is underway and long delivery items have been ordered. The plant will ultimately comprise four units with the first one to be completed in 1984 having a capacity of 37 000 Nm^3/hr of synthesis gas from 54 t/hr of lignite. Methanol (14 t/hr) will be produced from the synthesis gas. With subsequent expansions to a total of four gasification trains, the plant will have an ultimate output of 1×10^9 Nm^3/yr of synthesis gas from 2×10^6 t/yr of lignite, based on 8 000 hrs/year operation. The location of the new plant is advantageous to Rheinbraun in that the existing lignite preparation facilities, adequate supplies of steam and electricity, and a trained work force can be used.

3.3.4.2 Process Description

A block flowsheet of the pilot plant is shown in Figure 3.7.

The feedstock to the plant is dried lignite of moisture content 18%, which is usually used for briquetting. The lignite is stored and redried in a spiral tube pneumatic drier to a water content of between 8 and 12% depending on the test procedure. The lignite is brought to reaction pressure (1 000 - 1 100 kPa) in a lock hopper system. Simultaneously, limestone or lime is being fed into the gasifier by screw conveyor. In the gasifier, the lignite reacts with the preheated gasification agent, which is composed of either steam/oxygen or air, which is introduced at different levels. The product gas, laden with dust, leaves the gasifier at the top. The dust particles are removed in a primary cyclone and recycled to the gasifier. In a secondary cyclone the gas is further treated to remove minute particles which are disposed as fly ash. The gas is cooled in a waste heat boiler, generating superheated steam for use in the gasifier. Rheinbraun has the option of burning the hot raw gas in a nearby power station or further scrubbing the gas for feed to a test shift conversion unit to generate SNG. A sulfur resistant shift catalyst is required because of the sulfur compounds contained in the gas. Rheinbraun have carried out many tests with varying types of available catalysts and by varying quench conditions to obtain different steam-to-gas ratios.

3.3.4.3 Feedstock

The High-Temperature Winkler pilot plant has been demonstrated to process a wide range of coals, even those with a high ash content. The process can accept coal sizes from fines up to a recommended maximum size of 12 mm, as greater sizes tend to abrade the gasifier. Optimum moisture content of lignite is in the 8-12% range, though the unit has been demonstrated to operate at 18% moisture content.

An elemental analysis of a typical Rhenish brown coal processed in the test plant is as follows:

Size: 6-8 mm (59% less than 1 mm)

Component	Rhenish Brown Coal Weight Percent	Average Onakawana (m.a.f.)
C	68.1	73.04
H	5.2	4.69
O	25.2	19.66
N	0.5	0.86
S	0.5	1.75
Ash	4-5	18.80

3.3.4.4 Process Performance

The High-Temperature Winkler process is capable of producing a type of gas suitable for use as a raw gas for methanol synthesis. The optimum pressure for operation lies between 820 kPa and 1 030 kPa.

Pressurizing the gasifier decreases the number of units required for a given input. For example, seven gasifiers are required for a 140 kPa operation compared with four for 885 kPa in an oxygen-blown Winkler plant. Increasing pressure increases methane content in the synthesis gas, although this may be offset by raising the temperature.

Rheinbraun are performing further development work at raising the gasification pressure above the present 1 000 - 1 100 kPa, as this can reduce the plants energy requirement as well as increasing the specific throughput of the gasifier.

A carbon conversion rate of 96% and thermal efficiency of 82% have been achieved in the pilot plant.

The following table compares the results of lignite gasification with steam and oxygen in an atmospheric Winkler gasifier and the HTW pilot plant.

Gasifying Conditions		Atm. Winkler	H.T.W.
Temperature (°C)		950	990
Pressure (kPa)		120	1 000
Gasifying Agents			
Oxygen	Nm^3/kg dry lignite	.398	.38
Steam	kg/kg dry lignite	.167	.41
Specific Yield ($CO+H_2$)	Nm^3/tonne dry lignite	1 396	1 483
Specific Output ($CO+H_2$)	Nm^3/m^2 hr	2 122	5 004
Carbon Conversion %		91	96

These results show an increase in the specific synthesis gas yield due to the elevated temperature operation and partial recycling of the discharge dust and consequent increase in carbon conversion.

The specific synthesis gas production per m^2 and hour was more than doubled as a result of the pressure increase.

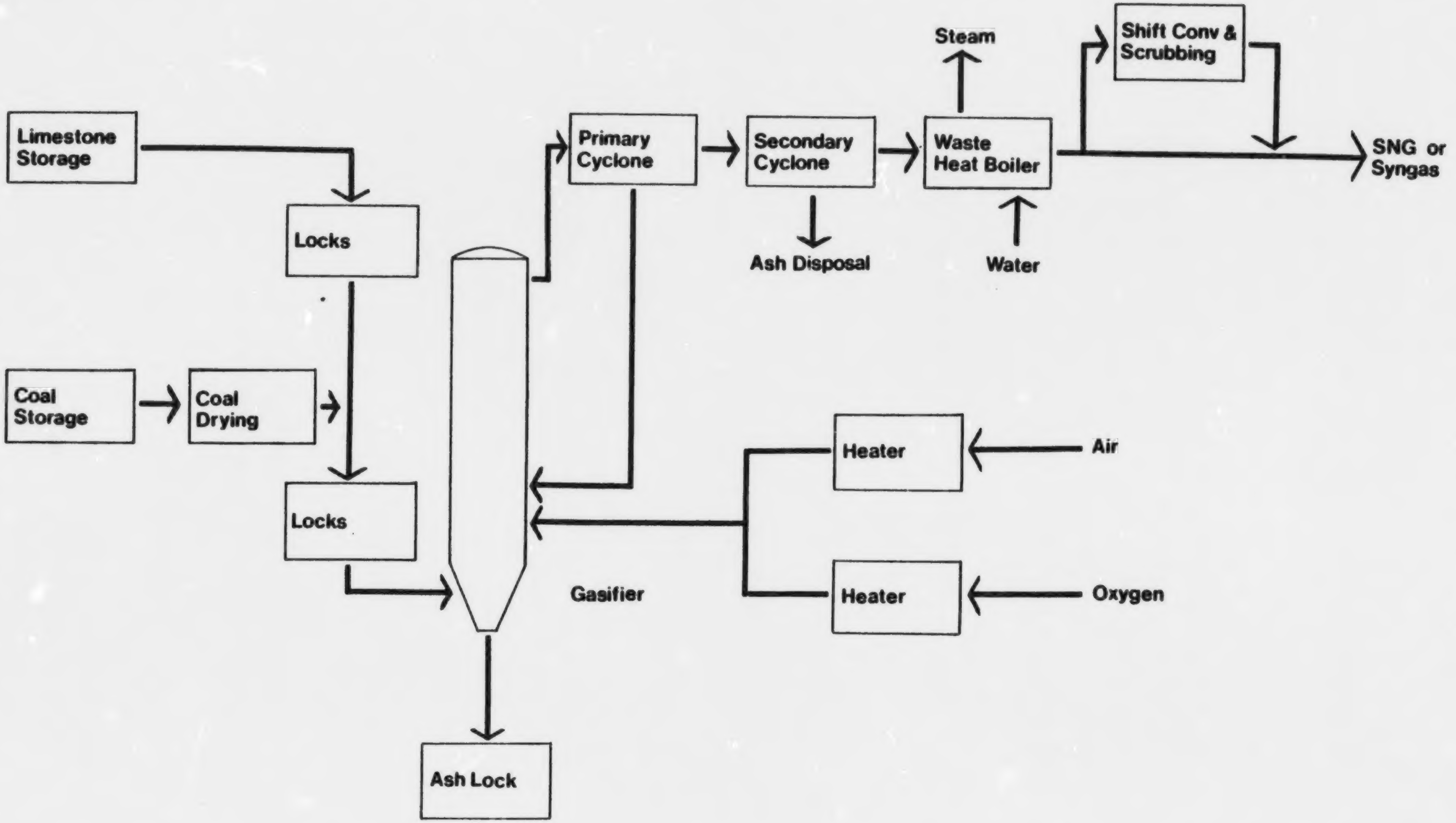
3.3.4.5 Operating Performance

The High-Temperature Winkler process has been reported to exhibit the following operating benefits:

- Easy start-up and shutdown procedures.
- Considerable flexibility to handle varying throughput rates.
- High temperature operation ensures that higher hydrocarbons are decomposed, thus ensuring that no tarry substances or liquid byproducts are produced.

Operating problems have however been reported in the performance of the cyclones in removal of fine particles entrained in the gas. Other problems encountered by Rheinbraun but stated as overcome include:

- Design and arrangement of gasifier nozzles.
- Lock-hopper malfunctioning.
- Fines return line malfunctioning.


Limestone Storage
Locks
Coal Storage
Coal Drying
Locks
Gasifier
Ash Lock
Primary Cyclone
Secondary Cyclone
Ash Disposal
Waste Heat Boiler
Steam
Water
Shift Conv & Scrubbing
SNG or Syngas
Heater
Air
Heater
Oxygen

High Temperature Winkler Process
Figure 3.7

3.3.5 TECHNOLOGY DEVELOPERS PREDICTIONS

3.3.5.1 Objectives

The rationale behind approaching the technology developers was to obtain their prediction on the performance of their process using Ontario and Onakawana lignites to produce methanol as end-product. Based on the assumption that lignites could be processed each developer was requested to provide information as follows:

- Recommended test programs to establish the design basis for a commercial plant, and costs and schedule related to these tests.
- Description of test facilities.
- Preliminary comments related to processing Ontario lignites such as predrying requirements, expected thermal efficiencies and carbon conversion.

The analyses (proximate, ultimate, ash), and ash fusion temperature of Ontario, Onakawana and Saskatchewan lignites were provided to assist each developer in the predictions.

3.3.5.2 Koppers-Totzek Process

.1 Process Developer

The process developer is GKT (Gesellschaft fur Kohle-Technologie) with head office in Essen, West Germany and U.S. contact address in Murray Hill, N.J.

.2 Test Programs and Cost

GKT has a bench-scale gasifier which processes approximately 6 kg/hr of dry feedstock. For large scale tests commercial units are available in facilities of GKT's clients. Bench-scale testing can be arranged at short notice and takes about 8 weeks whereas commercial gasification tests require 3 months. A bench scale test should provide sufficient data for the design basis for a commercial plant, unless the results indicate unusual characteristics of the lignite which would require additional large scale tests.

A laboratory test will provide a good basis for a computer analysis to deliver the data and information necessary to compile an information package suitable for a feasibility study. Costs provided are:

Laboratory Test	(5 kg sample)	\$ 9 750 U.S.
Bench Scale Test and Report	(300 kg sample)	\$135 000 U.S.
Commercial Test	(estimate only)	\$3.0 - \$4.0 million U.S.
Information package from lab test and computer analysis, depending on scope of battery limits		\$23 000 - \$80 000 U.S.

.3 Scope of Services

GKT does not sell a design package but provides complete engineering services within its battery limits. These battery limits include the gasification with wash water treatment and gas clean-up system (acid gas removal). In the case of methanol production, GKT can provide engineering for integrated units with synthesis gas as product delivered at battery limits.

.4 Process Suitability

Preliminary comments included:

- Lignite is a highly suitable feedstock for Koppers-Totzek process.

- GKT have not processed Canadian lignites, but does not expect significant deviation in gasifying Canadian lignites. However, these predictions can only be verified by laboratory testing.

- Carbon conversion is expected to be 98/99% with thermal efficiency in the 75% range (on a dry basis).

- Lignite has to be ground and dried as a preparation unit upstream of the gasifier. The recommended moisture content of feedstock is in the range of 6-8% by weight.

3.3.5.3 Lurgi Process

.1 Process Developer

The process developer is Lurgi Kohle and Mineraloltechnik GmbH, with head office in Frankfurt, West Germany and U.S. contact address in River Edge, New Jersey.

.2 Test Programs and Cost

For initial evaluation a bench scale laboratory test is sufficient to provide information for a preliminary plant design. Gasification parameters are estimated including gas composition, by-product generation, steam and oxygen requirements. Full scale tests can be performed at Sasol, South Africa or at the Ruhr-100 facility in Dorsten but these are not considered necessary for final plant design. Costs are

Lab-scale and interpretive report [50 lb sample]:	40 000 DM
Each additional sample	35 000 DM

Test work can be completed and report submitted within 8-10 weeks of receipt of the sample.

.3 Scope of Services

Lurgi technology is available for all the process on-sites including sulfur removal (Rectisol), methanol synthesis, methane reforming and waste water treatment.

.4 Process Suitability

Preliminary comments included:

- Typically, lignites are an ideal feedstock for Lurgi gasification exhibiting low oxygen requirements, and high reactivity. However, the high moisture content of the Ontario lignites (40-50%) is a major concern since predrying of the lignite to the maximum tolerable range of 30-35% may result in severe decrepitation. Significant fines generation could result which cannot be effectively processed in the Lurgi dry bottom gasifier. The bench scale test should indicate the magnitude of this problem. However, Lurgi has developed a circulating fluid bed gasification/combustion technology which could be considered as a more efficient alternative.

3.3.5.4 High-Temperature Winkler

.1 Process Developer

The process developer is Rheinische Braunkohlenwerke AG, (Rheinbraun) Cologne, West Germany. Their initial contact was in telex form and did not provide complete information in regard to test descriptions and tests required to prepare a preliminary design of commercial plant. Uhde Company provides engineering services for Rheinbraun.

.2 Test Program

Rheinbraun have available a PDU of capacity 35 kg/hr. The program would last 4 months which would require 4 tonnes of predried sample. The cost including a report is 290 000 DM (approximately $145 000 CAN). The program could also be broken down into three phases which were not described by Rheinbraun costing 35 000 DM, 30 000 DM and 225 000 DM ($17 500, $15 000, $112 500 CAN). Preliminary process design can be provided by Uhde for a complete plant at extra cost.

Large scale tests (1 t/hr) in the High Temperature Winkler pilot plant can be done to obtain further data for detailed design. This comprises physical and chemical laboratory tests, fluidization tests in a cold model and gasification tests including extended time tests for detailed mass, elemental and heat balances. Costs and sample sizes were not given.

.3 Process Suitability

Ontario lignite is considered a suitable feedstock for the High-Temperature Winkler process.

3.3.5.3 Westinghouse Process

.1 Process Developer

The process developer is Westinghouse Electric Corporation, Synthetics Fuels Division, Pittsburgh, Pa. The Canadian contact address is Westinghouse Canada, Calgary Alberta. A telex response was received which did not provide complete information to the inquiry package and this has been supplemented through verbal discussions.

.2 Test Program

Westinghouse has available laboratory bench scale facilities to perform the required analyses to provide data for conceptual design. The data can be used for heat and material balances, performance projecting, systems descriptions, equipment lists and cost estimates for a Westinghouse coal gasification system plant. The duration of the test program is approximately 12 weeks from receipt of sample to issue of report. A test sample, 50 kg (preferably wet) is sufficient, but it is recommended to repeat the entire analysis and preliminary design with a second sample.

.3 Scope of Services

Westinghouse battery limits are limited to lignite pressurization, gasification, particulate recovery, ash removal, raw gas heat recovery and gas quench systems. A raw gas composition within an accuracy of 1/2 of one percent will be provided.

.4 Process Suitability

The Westinghouse process is considered suitable to gasify Ontario lignites. It is recommended to pre-dry the lignite to 5-8 percent surface moisture but this would need confirmation when details of the steam and oxygen requirements of an integrated plant are better defined.

3.3.6 DOWNSTREAM UTILIZATION SUITABILITY

3.3.6.1 Downstream Utilization Potential

The product gas from each of the Lurgi, Westinghouse, High Temperature Winkler and Koppers-Totzek coal gasifiers is a raw synthesis gas containing H_2, CO, CO_2, CH_4, H_2S, and other minor component gases; the composition of this gas is a function of coal composition, type of gasifier used and operating conditions. This raw synthesis gas, with prior acid gas removal and H_2/CO ratio adjustment, can be used directly as a feedstock for further processing to a variety of products, such as methanol, synthetic gasoline via SASOL Fischer-Tropsch (F-T) or Mobil Methanol-to-Gasoline (MTG) technologies, ammonia, urea, hydrogen, glycols, acetic acid, polymethylen, vinyl acetate, etc.

Figure 3.8 illustrates the downstream utilization potential for the raw synthesis gas produced from the coal gasifiers. The required H_2/CO ratios for various product synthesis, to name a few, are:

- 2/1 for methanol synthesis
- 2/1 for SASOL F-T synthesis
- 2/1 for polymethylene synthesis
- 3/2 for glycol synthesis
- 2/2 for acetic acid synthesis.

For this study, downstream utilization of the raw synthesis gas for methanol production is emphasized.

3.3.6.2 Raw Synthesis Gas Composition

In terms of raw synthesis gas composition and processing requirements for downstream methanol production, the four gasifiers can be divided into three groups, that is:

(a) Lurgi gasifier
(b) High Temperature gasifiers: High Temperature Winkler and Koppers-Totzek.
(c) Westinghouse gasifier

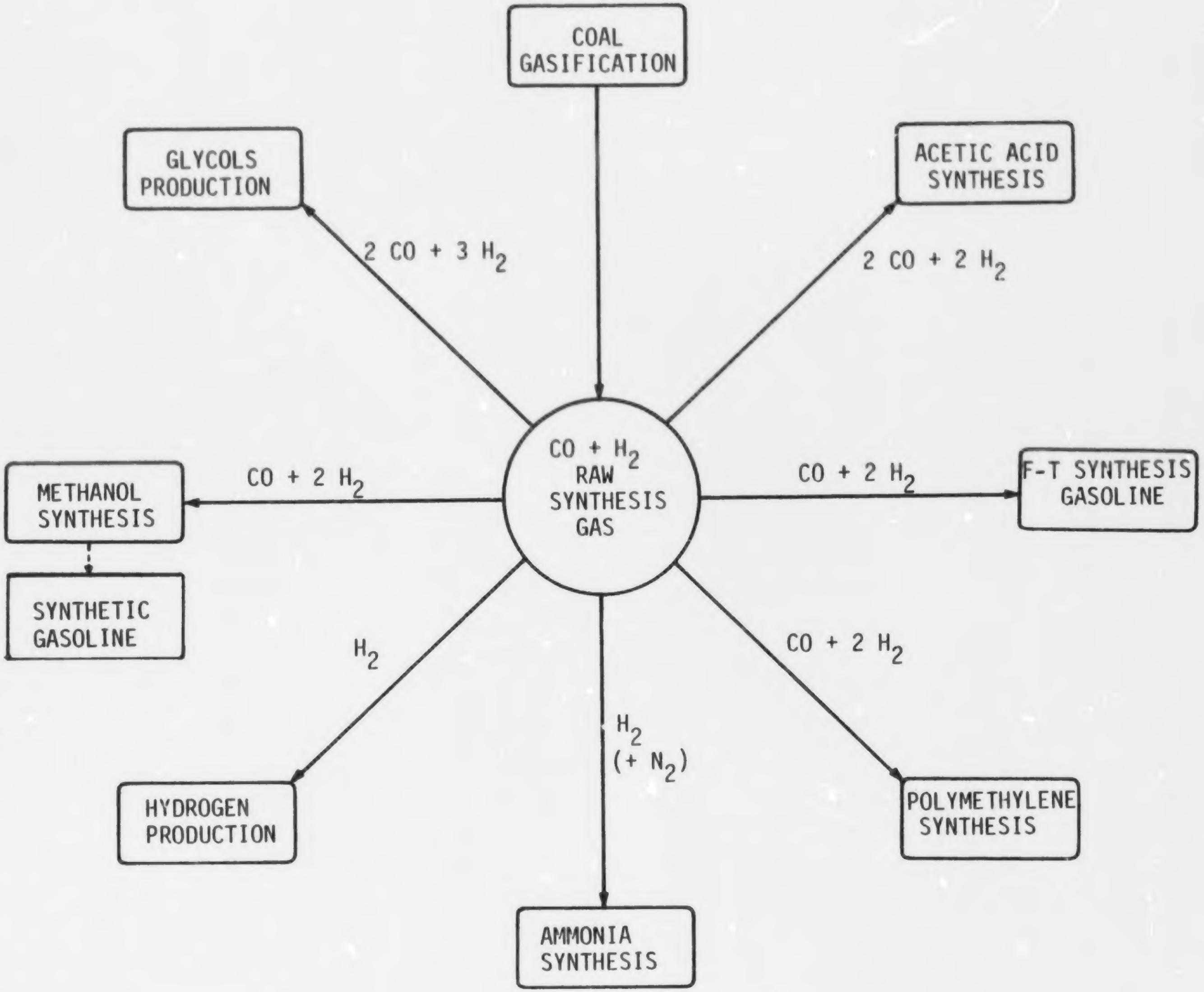

COAL GASIFICATION
GLYCOLS PRODUCTION
ACETIC ACID SYNTHESIS
2 CO + 3 H_2
2 CO + 2 H_2
CO + H_2 RAW SYNTHESIS GAS
METHANOL SYNTHESIS
CO + 2 H_2
CO + 2 H_2
F-T SYNTHESIS GASOLINE
SYNTHETIC GASOLINE
H_2
CO + 2 H_2
H_2 (+ N_2)
HYDROGEN PRODUCTION
POLYMETHYLENE SYNTHESIS
AMMONIA SYNTHESIS

FIGURE 3.8 DOWNSTREAM UTILIZATION OF RAW SYNTHESIS GAS FROM COAL GASIFICATION

The raw synthesis gas from the Lurgi gasifier, while processing a lignite of the Onakawana type, is expected to have a H_2/CO ratio of about 2.3/1. The raw synthesis gas also contains a high CH_4 content (about 15%) and small amounts of $C_2H_4 + C_2H_6$, in addition to CO_2, H_2S and COS. For the two high-temperature gasifiers, the raw synthesis gas is likely to have a H_2/CO ratio of about 0.5/1 - 0.8/1, while the methane content is relatively low.

The raw synthesis gas from the Westinghouse gasifier, containing an expected H_2/CO ratio of 0.6/1 and a methane content of about 5%, is intermediate between the Lurgi synthesis gas (about 15% methane) and the low methane content synthesis gas from the H.T. Winkler or Koppers-Totzek gasifiers.

Because of the distinct difference between these three types of raw synthesis gas, each type has its unique downstream processing requirements for methanol production. These requirements and the alternatives are discussed in the following section.

3.3.6.3 Downstream Processing Requirements

In general, the downstream processing requirements for the raw synthesis gas to produce methanol are:

- removal of H_2S, COS and most of the CO_2
- adjustment of H_2/CO ratio to 2/1
- compression
- methanol synthesis
- methanol purification.

The process steps required to adjust the H_2/CO ratio are dependent upon the following factors:

- raw synthesis gas composition
- availability and economics of supplementary H_2-rich synthesis gas or hydrogen
- optimum stiochiometric relationship between the raw synthesis gas and the supplementary synthesis gas or hydrogen.

The compression requirement for the synthesis gas is:

- extensive for the Koppers-Totzek gasifier (raw synthesis gas from near atmospheric pressure to 5-6 MPa)
- normal for the other three gasifiers (from 2-3 MPa to 5-6 MPa).

Based on the three groups of gasifiers, the schematics and alternatives for downstream processing are discussed as follows:

Lurgi Gasifier

Base Case

Figure 3.9 illustrates the base case of downstream processing requirements for methanol production, using the raw synthesis gas from Lurgi gasifiers as a feedstock.

The raw synthesis gas is first processed in an acid gas removal unit (Rectisol), using methanol as physical solvent to remove H_2S and COS, CO_2 and naphtha. The acid gas stream is sent to a sulphur recovery unit (Stretford) to recover the elemental sulphur as a by-product. The naphtha from the Rectisol unit is used as a supplementary fuel for steam generation or is sold as a byproduct.

The purified synthesis gas, containing about 56% H_2, 23% CO, 5% CO_2, 15% CH_4, 0.5% C_2H_4 + C_2H_6, and traces of other gases, has a H_2/CO ratio of about 2.4/1. This gas can be used directly as a feedstock for methanol synthesis, using a copper-based catalyst.

$$CO + 2H_2 \longrightarrow CH_3OH$$

While the purified synthesis gas contains a high content of CH_4, this methane remains inert in the methanol synthesis converter and exits as a purge gas. This purge gas is then sent to a steam reforming unit. In the presence of nickel catalyst, steam reacts with methane at elevated temperature and pressure to give a reformed gas consisting of H_2, CO, CO_2 and CH_4; H_2/CO in the range of 3:1.

This reformed gas is combined with the coal-derived synthesis gas, compressed from 2-3 MPa to 5-6 MPa and fed to the methanol synthesis converter. The crude methanol from the converter is then sent to a distillation unit to obtain fuel grade methanol, containing 99.9% methanol and higher alcohols and 0.1% water (maximum).

Purge gas from the methanol purification unit is used as a supplementary fuel gas for the steam reforming system.

Alternative

An alternate scheme for the downstream processing of raw synthesis gas from the Lurgi gasifier is illutrated in Figure 3.10. This scheme uses catalytic partial oxidation, instead of steam reforming, to convert methane in the methanol converter purge gas into a H_2-rich synthesis gas.

The methane-containing purge gas, along with oxygen and steam, are fed to a catalytic partial oxidation unit. Reactions in this unit occur in two zones. The methane is fed into the top zone where it is partially combusted by the oxygen. The exothermic combustion reaction supplies the heat necessary for the subsequent endothermic reforming reaction. The reforming reaction occurs in the second reaction zone, which consists of a fixed bed catalyst. The resultant synthesis gas has a H_2/CO ratio of 2/1.

High Temperature Gasifiers

The raw synthesis gas from the high-temperature gasifiers has a typical H_2/CO ratio of 1:2, in addition to other gas components such as CO_2, H_2S, COS, etc. To adjust the H_2/CO ratio to 2:1 for methanol production, high-temperature CO shift conversion or supplementary hydrogen source is required.

Base Case:

Figure 3.11 illustrates the base case of downstream processing requirements for the high temperature gasifiers. In the base case, the raw synthesis gas from the coal gasifier is first processed in a high temperature CO shift converter to produce a synthesis gas with an H_2/CO ratio of 2/1. This synthesis gas is then sent to an acid gas removal (Rectisol) unit to remove CO_2, H_2S and COS. The recovered sulphides are sent to a sulphur recovery (Stretford) unit to recover the elemental sulphur as a by-product.

The purified synthesis gas is compressed to 5-6 MPa and than sent to a methanol synthesis unit. The crude methanol produced is further purified in a distillation unit to obtain the fuel grade methanol.

The purge gases from the methanol synthesis unit and the methanol purification unit are used as supplementary fuel for steam generation.

Alternative 1:

Figure 3.12 illustrates Alternative 1 for the downstream processing of raw synthesis gas from the high temperature gasifiers.

The raw synthesis gas is first processed in an acid gas removal (Rectisol) unit to remove CO_2, H_2S and COS. The sulphides are sent to a sulphur recovery (Stretford) unit to recover the elemental sulphur as a by-product.

The purified synthesis gas has a H_2/CO of 0.5/1 to 0.8/1 and is hydrogen-deficient for methanol production. The adjustment of H_2/CO ratio to 2/1 requires mixing of this purified synthesis gas with H_2-rich synthesis gas, which is produced by steam reforming of natural gas.

The reformed synthesis gas has a H_2/CO of about 3:1. This gas is combined with the purified gasifier synthesis gas to achieve a H_2/CO ratio of 2/1. The combined gas is compressed to about 5-6 MPa prior to being sent to a methanol synthesis unit. The crude methanol produced is then purified in a distillation unit to obtain the fuel grade methanol.

The purge gases from the methanol converter and the distillation unit are sent to the steam reforming unit as supplementary fuel.

Alternative 2:

Figure 3.13 illustrates Alternative 2 for the downstream processing of raw synthesis gas.

Alternative 2 is the same as Alternative 1 in terms of the process steps for acid gas removal, sulphur recovery, methanol synthesis and methanol purification. However, the adjustment of H_2/CO ratio is achieved by the addition of hydrogen, which is produced by processing the natural gas in a steam reforming/shift conversion/PSA purification system.

Natural gas is desulphurized and steam reformed to produce a synthesis gas with a H_2/CO ratio of 3/1. The reformed gas is sent to a high temperature CO shift converter, where CO reacts with steam to yield additional hydrogen and CO_2. The high temperature shift effluent is cooled to ambient temperature and enters the PSA unit. The resultant product, hydrogen at 2.4 MPa, has a purity of 99.9%. This hydrogen is then combined with the purified synthesis gas from the coal gasifier to produce a synthesis gas with an H_2/CO ratio of 2/1 for methanol production.

The purge gas from the PSA unit, along with the purge gases from the methanol converter and the purification unit, are accumulated in the purge gas drum and are used as fuel in the reforming furnace.

Alternative 3:

Figure 3.14 illustrates Alternative 3 for the downstream processing of raw synthesis gas from the high-temperature gasifiers.

Alternative 3 is the same as Alternative 2 in terms of the process steps for acid gas removal, sulphur recovery, methanol synthesis and methanol purification, and also the use of natural gas originated hydrogen to adjust the H_2/CO ratio of synthesis gas. However, there are a number of differences between these two alternatives:

(a) A conventional steam reforming process is used in Alternative 3 to produce hydrogen from natural gas; this process consists of desulphurization, steam reforming, CO shift, and CO_2 removal units.

(b) In Alternative 3 the raw synthesis gas from the coal gasifier and the natural gas-derived syngas from the CO shift converter share a common CO_2 removal unit, which is a part of the Rectisol acid gas removal system.

The product gas from the steam reforming plant contains 97-98% H_2. The impurities in the product hydrogen are predominately CO, CO_2 and methane.

This hydrogen gas is combined with the purified coal gasifier synthesis gas to obtain a methanol synthesis gas with an H_2/CO ratio of 2/1.

The purge gases from the methanol synthesis converter and the methanol purification unit are used in the steam reforming furnace as supplementary fuel.

Alternative 4:

Figure 3.15 illustrates Alternative 4 for the downstream processing requirements for the high temperature gasifiers.

Alternative 4 is the same as Alternative 2 in terms of the process steps for acid gas removal, sulphur recovery, methanol synthesis and purification, and also the use of hydrogen to adjust H_2/CO ratio of synthesis gas.

The required hydrogen is supplied by the electrolysis of water, which yields high purity hydrogen and oxygen. The electrolytic oxygen is utilized in the gasifiers for gasification of the coal. The electrolytic hydrogen is combined with the purified gasifier synthesis gas to obtain the methanol synthesis gas with an H_2/CO ratio of 2/1.

The purge gases from the methanol synthesis converter and the methanol purification unit are accumulated in the purge gas drum and used as fuel for steam generation.

Westinghouse Gasifier

As the raw synthesis gas from the Westinghouse gasifier requires adjustment of the H_2/CO ratio and as it contains a relatively high methane content, optimum processing will probably be a combination of the alternatives discussed for the Lurgi and High Temperature gasifiers. Effective utilization of the methane content of the synthesis gas will be required; otherwise overall thermal efficiency will be lower than anticipated.

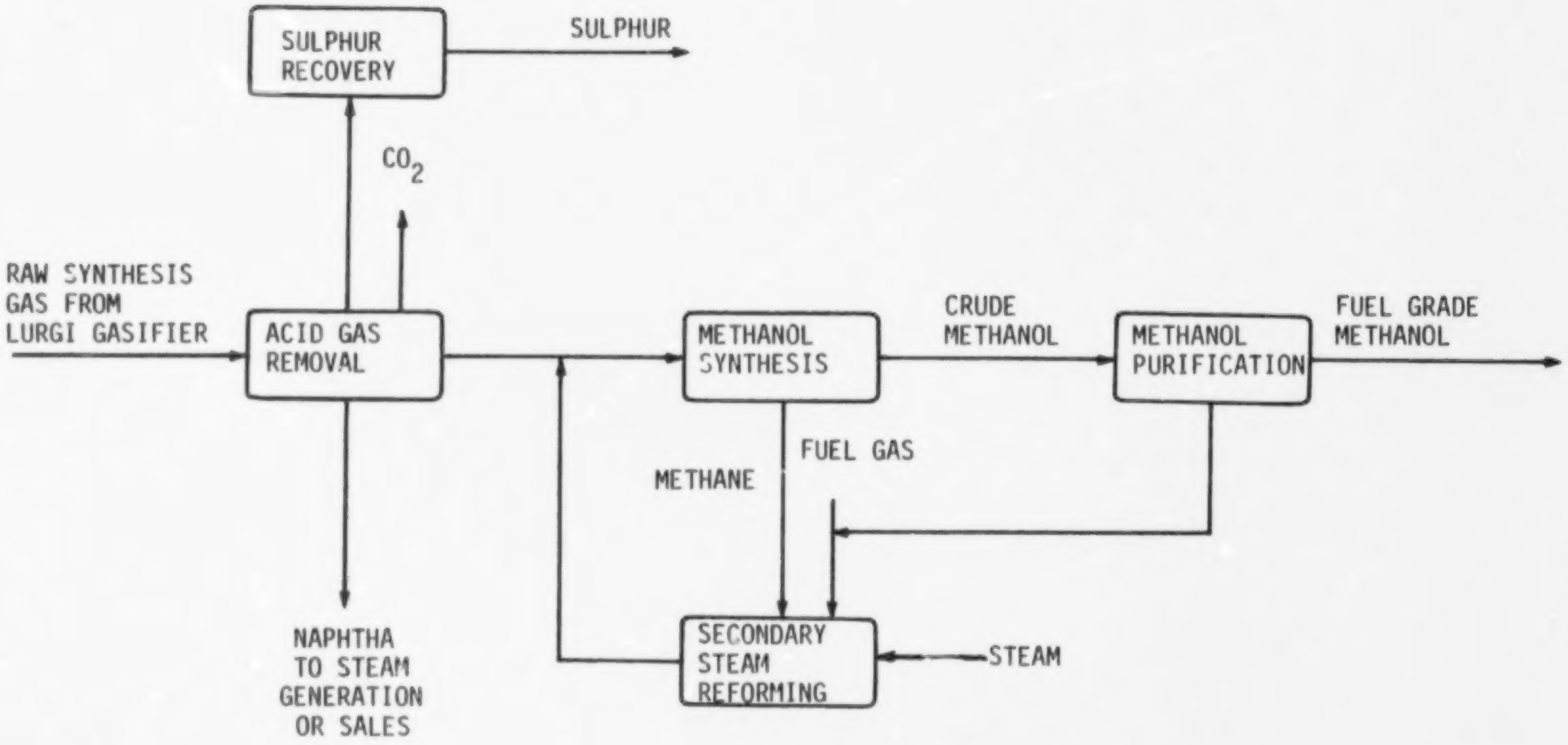

SULPHUR RECOVERY
SULPHUR
CO_2
RAW SYNTHESIS GAS FROM LURGI GASIFIER
ACID GAS REMOVAL
METHANOL SYNTHESIS
CRUDE METHANOL
METHANOL PURIFICATION
FUEL GRADE METHANOL
FUEL GAS
METHANE
NAPHTHA TO STEAM GENERATION OR SALES
SECONDARY STEAM REFORMING
STEAM

FIGURE 3.9 METHANOL PRODUCTION FROM LURGI GASIFIER RAW SYNTHESIS GAS

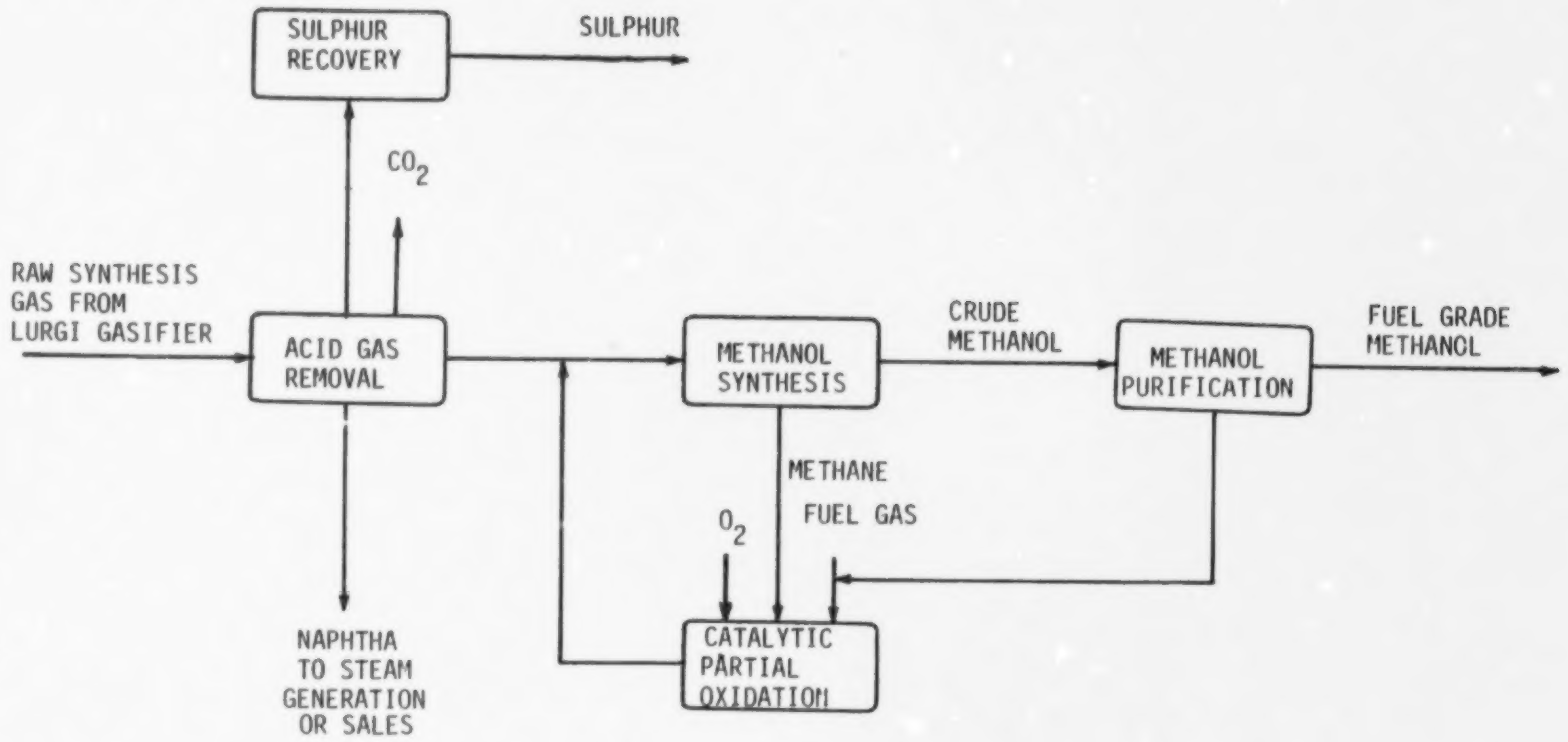

SULPHUR RECOVERY
SULPHUR
CO_2
RAW SYNTHESIS GAS FROM LURGI GASIFIER
ACID GAS REMOVAL
METHANOL SYNTHESIS
CRUDE METHANOL
METHANOL PURIFICATION
FUEL GRADE METHANOL
METHANE
O_2
FUEL GAS
NAPHTHA TO STEAM GENERATION OR SALES
CATALYTIC PARTIAL OXIDATION

FIGURE 3.10 ALTERNATE PROCESS FOR METHANOL PRODUCTION FROM LURGI GASIFIER RAW SYNTHESIS GAS

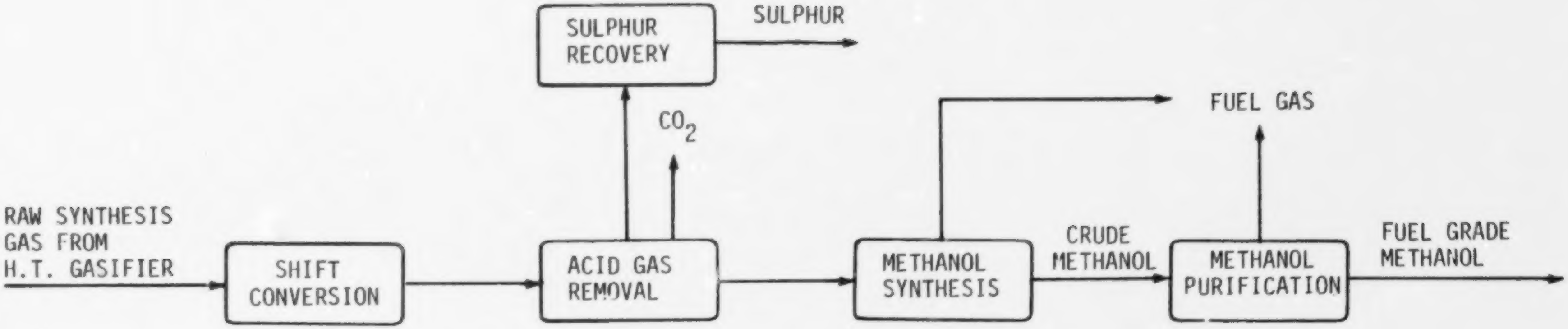

RAW SYNTHESIS GAS FROM H.T. GASIFIER
SHIFT CONVERSION
ACID GAS REMOVAL
SULPHUR RECOVERY
SULPHUR
CO_2
METHANOL SYNTHESIS
FUEL GAS
CRUDE METHANOL
METHANOL PURIFICATION
FUEL GRADE METHANOL

FIGURE 3.11 METHANOL PRODUCTION FROM HIGH TEMPERATURE COAL GASIFIER RAW SYNTHESIS GAS - BASE CASE

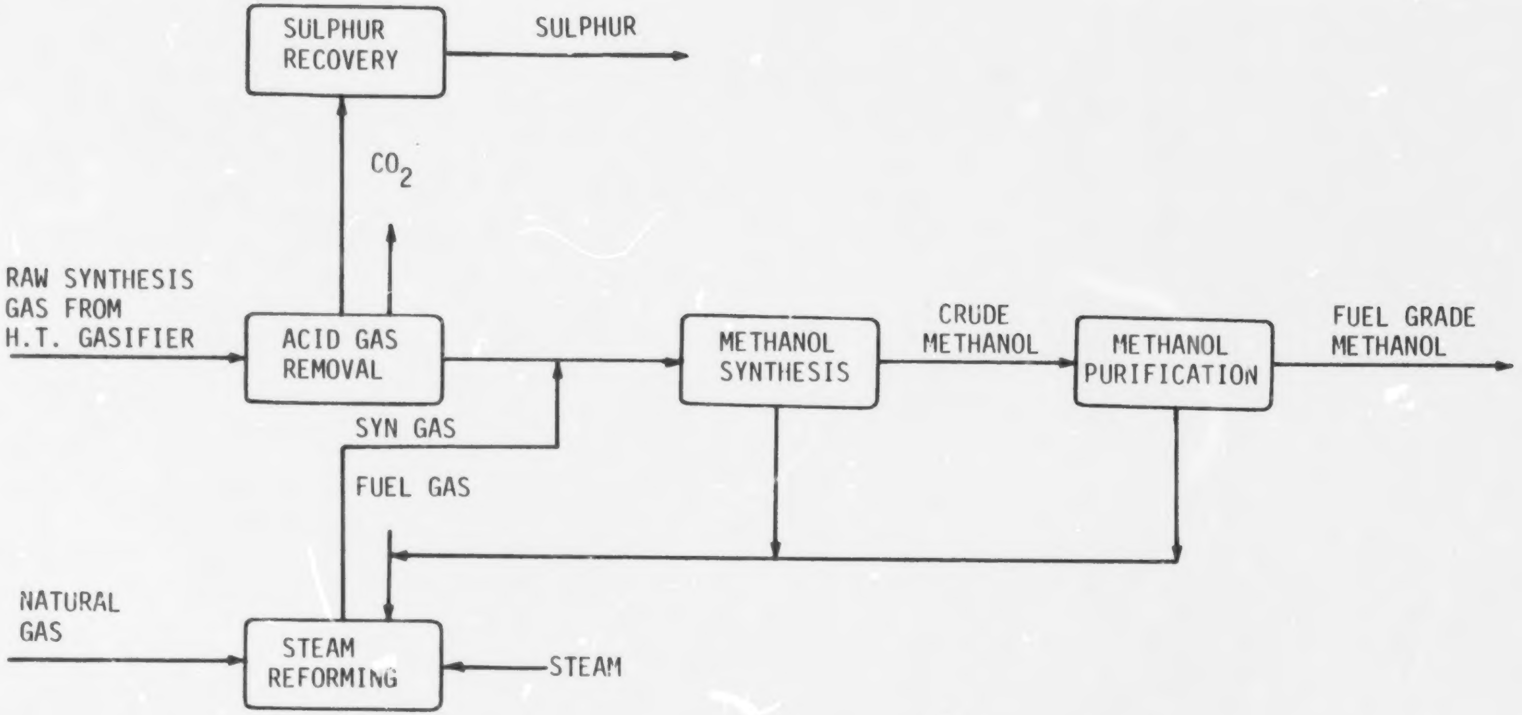

SULPHUR
RECOVERY
SULPHUR
CO_2
RAW SYNTHESIS
GAS FROM
H.T. GASIFIER
ACID GAS
REMOVAL
METHANOL
SYNTHESIS
CRUDE
METHANOL
METHANOL
PURIFICATION
FUEL GRADE
METHANOL
SYN GAS
FUEL GAS
NATURAL
GAS
STEAM
REFORMING
STEAM

FIGURE 3.12 METHANOL PRODUCTION FROM HIGH TEMPERATURE COAL GASIFIER RAW SYNTHESIS GAS - ALTERNATIVE 1

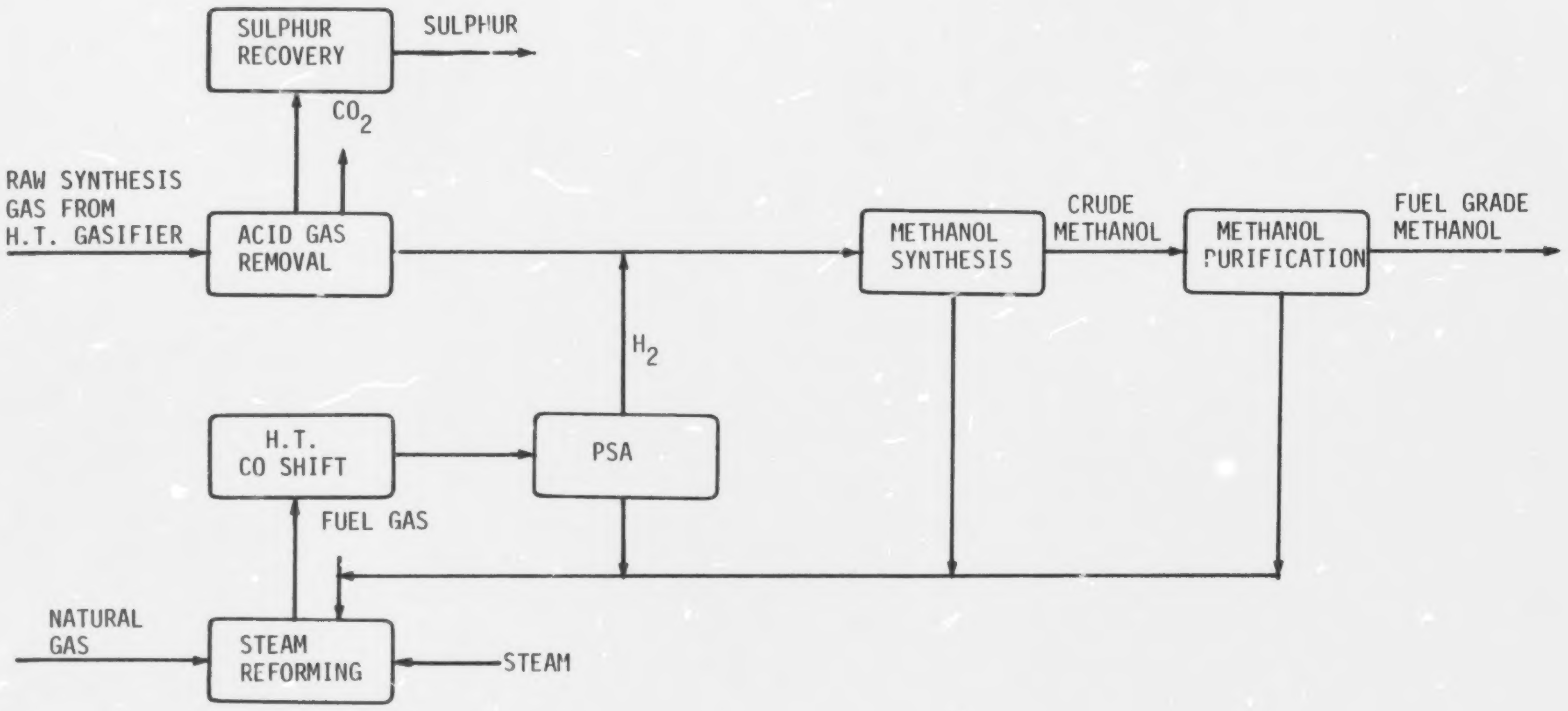

SULPHUR RECOVERY
SULPHUR
CO_2
RAW SYNTHESIS GAS FROM H.T. GASIFIER
ACID GAS REMOVAL
METHANOL SYNTHESIS
CRUDE METHANOL
METHANOL PURIFICATION
FUEL GRADE METHANOL
H_2
H.T. CO SHIFT
PSA
FUEL GAS
NATURAL GAS
STEAM REFORMING
STEAM

FIGURE 3.13 METHANOL PRODUCTION FROM HIGH TEMPERATURE COAL GASIFIER RAW SYNTHESIS GAS - ALTERNATIVE 2

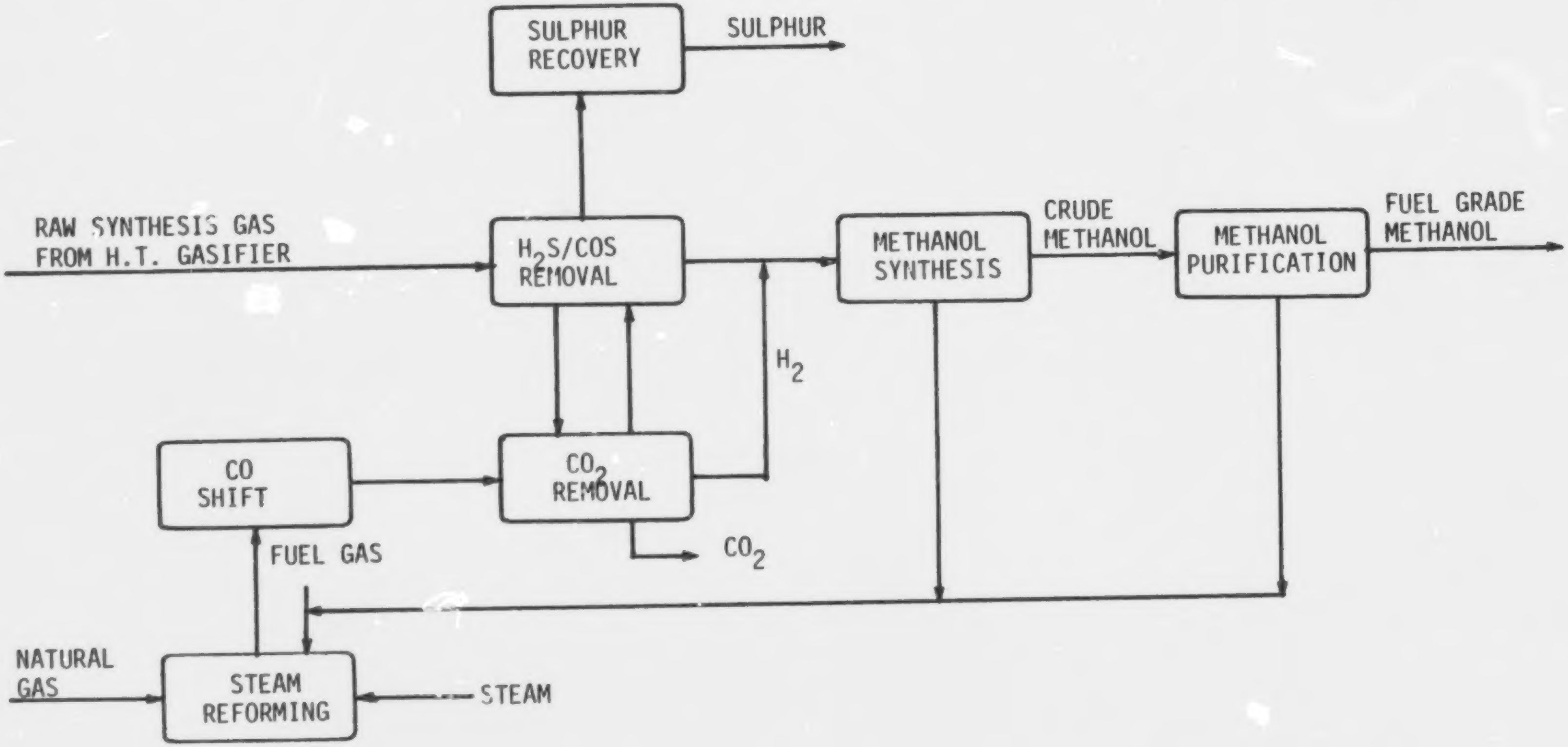

SULPHUR RECOVERY
SULPHUR
RAW SYNTHESIS GAS FROM H.T. GASIFIER
H_2S/COS REMOVAL
METHANOL SYNTHESIS
CRUDE METHANOL
METHANOL PURIFICATION
FUEL GRADE METHANOL
H_2
CO SHIFT
CO_2 REMOVAL
CO_2
FUEL GAS
NATURAL GAS
STEAM REFORMING
STEAM

FIGURE 3.14 METHANOL PRODUCTION FROM HIGH TEMPERATURE GASIFIER RAW SYNTHESIS GAS - ALTERNATIVE 3

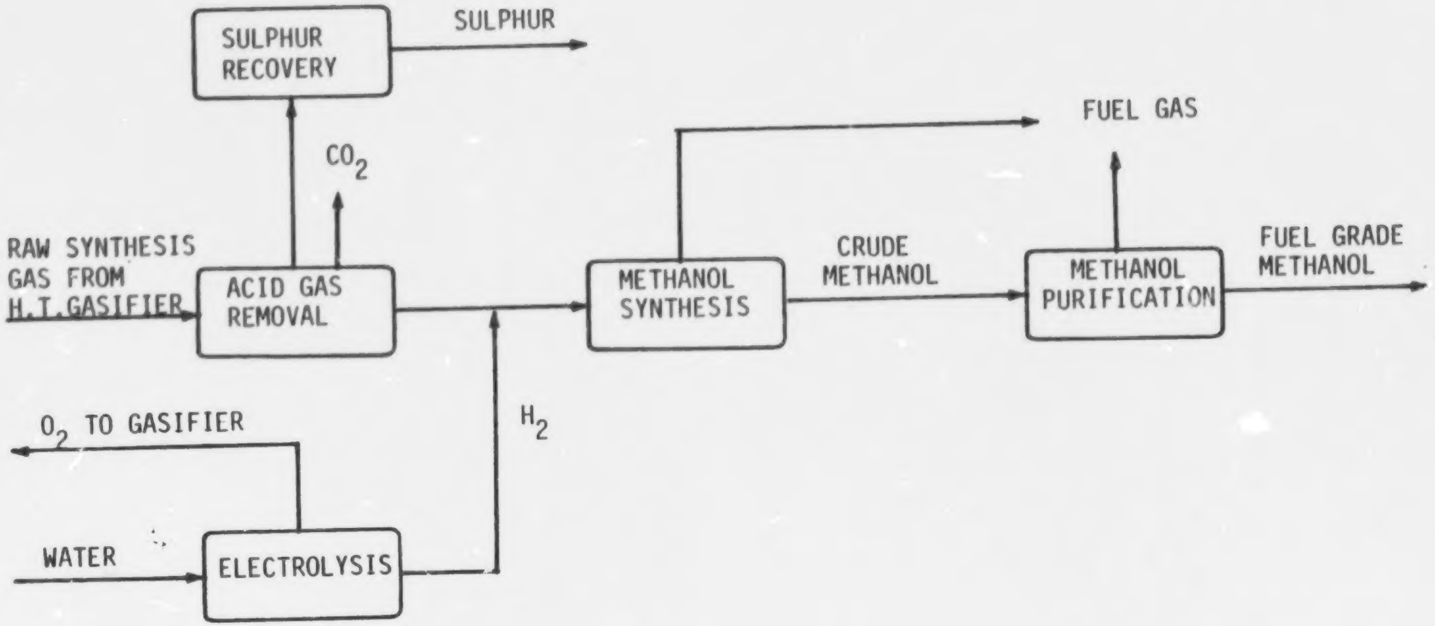

SULPHUR RECOVERY
SULPHUR
FUEL GAS
CO_2
RAW SYNTHESIS GAS FROM H.T.GASIFIER
ACID GAS REMOVAL
METHANOL SYNTHESIS
CRUDE METHANOL
METHANOL PURIFICATION
FUEL GRADE METHANOL
O_2 TO GASIFIER
H_2
WATER
ELECTROLYSIS

FIGURE 3.15 METHANOL PRODUCTION FROM HIGH TEMPERATURE GASIFIER RAW SYNTHESIS GAS - ALTERNATIVE 4

3.3.7 ENVIRONMENTAL CONSIDERATIONS

3.3.7.1 Waste Characteristics

The sources and characteristics of wastewater, gaseous effluent and solid waste from a typical coal gasification plant are identified by process unit and are given in Tables 3.5, 3.6 and 3.7.

The composition of each of the waste streams is a function of coal composition, type of gasifier used and operating conditions. The contaminants as well as their concentrations in the waste streams, therefore, vary from one coal gasification process to another.

Among the four coal gasification processes discussed previously, the Lurgi gasifier operates at a low temperature, resulting in a wider range of by-products and, therefore, a larger amount of contaminants in the wastewater. Also, because of the extensive use of steam to prevent the ash slagging, Lurgi gasifiers generate a larger volume of wastewater.

The gaseous effluent characteristics for the four coal gasification processes are similar in terms of quality and quantity. The product gases from the gasifiers contain particulates, sulphides and carbon dioxide, which are removed in a series of treatment systems to produce clean synthesis gas suitable for downstream utilization. Other process areas generate dust and small amounts of volatile organics and inorganics, which are common for all gasification processes.

The solid wastes from the four gasification processes are also similar in their characteristics. They include the ash or slag generated by the gasifiers, the rejects from the coal crushing and sizing areas, the sludges from the wastewater treatment plants, and the particulate matters collected from the particulate removal systems.

The comparative waste characteristics for four coal gasification processes are listed in Table 3.8 and are highlighted in the following paragraphs.

Wastewater

All four processes generate wastewater containing suspended solids, sulphides, ammonia, phenols, cyanide and thiocyanate. However, the contents of phenol, cyanide and thiocyanate in the wastewater of the Koppers-Totzek, Westinghouse and High Temperature Winkler processes are relatively small compared to that of the Lurgi process. Also, tar oils are present only in the Lurgi process wastewater.

The total volume of the wastewater generated from the Lurgi gasifier is notably larger than that from the other three coal gasification processes.

Gaseous Effluent

The gaseous effluents from all four coal gasification processes contain the same types of contaminants: suspended particulate matters, hydrogen sulphide, COS and trace amount of volatile organics and inorganics. During the start-up operation, the gasifier may generate tar and oil aerosols, which require proper control. The fugitive emissions from the gasifiers may contain hazardous species that are present in the raw synthesis gas, such as hydrogen cyanide.

Solid Waste

Because of the high operating temperature and high carbon conversion, the Koppers-Totzek process produces a slag which is practically inert. The fluidized bed gasifiers, Westinghouse and High Temperature Winkler, have a lower carbon conversion, resulting in a higher carbon content in the ash. The ash from the low temperature Lurgi gasifier contains unreacted carbon and a small amount of heavy hydrocarbons.

For all four of the gasification processes, the coal preparation areas generate solid wastes containing mineral matter rejects. Other solid wastes include the sludges produced from the filtration, flocculation and biological oxidation for wastewater treatment, and the particulate matter collected from the particulate removal systems.

3.3.7.2 Environmental Regulations

Prior to their discharge or disposal to the environment, all wastes from a coal gasification complex are to be treated to the acceptable limits set by the regulatory agencies. Environmental control measures, including treatments of gaseous and liquid effluents and solid waste management, therefore, are to be based on the best available control technology in order to satisfy the following Ontario and Federal regulatory requirements:

Ontario

- Environmental Protection Act	R.S.O. 1980, c. 141
- Air Pollution Control Regulations	R.R.O. 1980, Reg. 308
- Ambient Air Quality Criteria Regulations	R.R.O. 1980, Reg. 296
- Guidelines and Criteria for Water Quality Management in Ontario	Ministry of Environment, 1973
- Environmental Protection (General - Waste Management) Regulations	R.R.O. 1980, Reg. 309
- Public Health Act	R.R.O. 1980, c. 409

Federal

- Ambient Air Quality Objectives, No. 1	CRC, Vol. IV, c. 403
- Ambient Air Quality Objectives, No. 2	CRC, Vol. IV, c. 404
- Ambient Air Quality Objectives No. 3	SOR/78-74
- Petroleum Refinery Liquid Effluent Regulations	CRC, Vol. VII, c. 828
- Petroleum Refinery Liquid Effluent Guidelines (Acute Toxicity of Liquid)	January 1974.

3.3.7.3 Treatment Requirements

This section describes the waste treatment requirements for a coal gasification complex based on the four gasification processes. Applying the best available control technology for waste treatment and disposal, it appears that there is no problem for any of the four coal gasification processes to meet environmental regulatory criteria.

The treatment requirements are discussed below according to the type of waste.

Wastewater Treatment

Depending upon the types of contaminant to be removed from the wastewater, the treatment requirements may include oil separation, filtration, flocculation, solvent extraction, activated carbon adsorption and biological oxidation. The selection of the treatment process is determined by the following factors: contaminant removal efficiency, utility requirements and sequential treatment or disposal requirements.

Table 3.9 lists the specific wastewater treatment requirements and types of contaminant to be removed for each of the four coal gasification processes.

In summary, the Westinghouse, High Temperature Winkler and Koppers-Totzek processes have similar treatment requirements: filtration/ flocculation, sour gas stripping and biological oxidation. By comparison, the Lurgi gasification process requires the most extensive wastewater treatment, i.e. (a) larger volume of wastewater to be treated, and (b) additional oil-water separation and solvent extraction (Pheno-solvan) units are required.

Gaseous Effluent Treatment

The major gaseous effluent treatment requirements are the same for all of the four coal gasification processes, i.e. particulate removal and acid gas removal.

Particulate removal from the gasifier product gas is generally accomplished through the use of dry collectors such as cyclones or electrostatic precipitators and/or the use of water or oil scrubbers. Scrubbers are normally employed when tars and oils, as well as other impurities such as ammonia, must also be removed from the product gas, as in the case of the Lurgi gasifier. Moreover, the use of water or oil scrubbers cools the product gas. Other candidate cooling processes include waste heat boilers and air or water-cooled heat exchangers.

Acid gas removal processes remove compounds such as H_2S, COS, HCN and CO_2 from the product gas. In theory, either high temperature (>150°C) or low temperature acid gas removal processes can be used. High temperature processes utilize such sorbents as iron oxide, dolomites, and molten metals. They may yield higher energy efficiencies, but are only in the early development stage. Thus, most existing and proposed gasification plants use low temperature processes.

A variety of low-temperature acid gas removal processes are commercially available. Most processes were developed for use in the natural gas, refinery, and chemical process industries. There do not appear to be any insurmountable problems associated with their use in coal gasification plants. The following acid gas removal processes are the most likely candidates for near-term commercial use in coal gasification systems:

- Physical solvent processes: Rectisol, Selexol, Purisol, Estasolvan and Fluor Solvent.
- Chemical solvent processes: MEA, MDEA, DEA, DIPA, DGA and Benfield.
- Combination chemical/physical solvent processes: Amisol and Sulfinol.
- Direct conversion process: Stretford.

Among the above 14 options, the Rectisol unit, followed by a Stretford unit to recover the elemental sulphur, can be applied to all of the four coal gasification processes for acid gas removal.

The Rectisol process is used by SASOL in South Africa for acid gas removal from Lurgi gasifier-produced synthesis gas.

Solid Waste Management

The solid wastes from all four coal gasification processes can be disposed of in the same manner. The gasifier ash/slag, wastewater treatment sludges, collected particulate matters and coal crushing/sizing rejects can be transported to suitable sites for landfilling. Depending upon the specific characteristics of the solid wastes, lining and containment structures may be required to prevent any leaching of contaminants and subsequent seepage into nearby groundwater or surface water systems.

TABLE 3.5

COAL GASIFICATION PLANT WASTEWATER SOURCES AND CHARACTERISTICS

PROCESS UNIT	WASTEWATER SOURCE	POSSIBLE CONTAMINANTS
Coal Preparation and Storage	Coal-pile runoff; coal crushing/cleaning wastes	Suspended solids; dissolved Organics and inorganics
Gasifier (wet ash)	Ash quench/sluice water	Suspended solids; dissolved inorganics
Particulate Removal; Gas Quenching and Cooling; Acid Gas Removal	Process condensate; unrecoverable solvent	Suspended solids; non-emulsified oils, dissolved organics and inorganics; spent solvent
Gas/Liquor Separator	Gas liquor	Suspended solids; dissolved organics and inorganics
Cooling Tower	Blowdown	Suspended solids; dissolved organics and inorganics
Utility System	Blowdown	Dissolved inorganics; suspended solids
Wastewater Treatment	Sludges	Semi-solids

TABLE 3.6

COAL GASIFICATION PLANT GASEOUS EFFLUENT SOURCES AND CHARACTERISTICS

PROCESS UNIT	GASEOUS EFFLUENT SOURCE	POSSIBLE CONTAMINANTS
Coal Preparation and Storage	Coal pile dust emissions; coal crushing/sizing dust; coal drying dust and combustion	Dust; combustion gases; organics
Gasifier	Start-up vent stream; fugitive emissions	Tar and oil aerosols; sulphur compounds
Ash Removal	Vent gases	Volatile organics and inorganics
Acid Gas Removal	Tail gases	Trace Sulphur compounds; trace hydrocarbons
Boiler Plant	Flue gas	Sulphur oxides

TABLE 3.7

COAL GASIFICATION SOLID WASTE SOURCE AND CHARACTERISTICS

PROCESS UNIT	SOLID WASTE SOURCE	POSSIBLE CONTAMINANTS
Coal Preparation	Solid wastes from crushing and sizing	Mineral matter rejects
Particulate Removal	Collected particulate matters	Unreacted carbon; sulphur compounds; organics
Ash Removal	Ash or slag	Hydrocarbons
Wastewater Treatment	Sludges	-

TABLE 3.8

WASTE CHARACTERISTICS OF COAL GASIFICATION PROCESSES

WASTE / PROCESS	LURGI*	WESTINGHOUSE	HTW	K-T
Wastewater				
Suspended solids	X	X	X	X
Phenol	X	min.	min.	min.
Cyanide & Thiocyanate	X	min.	min.	min.
Ammonia	X	X	X	X
Tar Oils	X	-	-	-
Sulphides	X	X	X	X
Gaseous Effluent				
Particulate Matters	X	X	X	X
H_2S & COS	X	X	X	X
Solid Waste				
Ash	X	X	X	Inert slag
Sludges and rejects	X	X	X	X

* Lurgi gasifier produces a larger volume of wastewater.

TABLE 3.9

WASTEWATER TREATMENT REQUIREMENTS FOR COAL GASIFICATION PROCESSES

TREATMENT GASIFICATION	LURGI	WESTINGHOUSE	HTW	K-T	CONTAMINANTS REMOVED
Filtration; Flocculation	X	X	X	X	Suspended solids
Oil Separation	X	-	-	-	Tar oils
Solvent Extraction (Phenosolvan)	X	-	-	-	Phenol
Sour Gas Stripping	X	X	X	X	Sulphide, ammonia
Biological Oxidation	X	X	X	X	Dissolved organics, cyanide and thiocyanate, sulphide

Kilborn has followed standard professional procedures in preparing the contents of this report, but it is based in part on details, information and assumptions provided by others and Kilborn, therefore, cannot guarantee the correctness of the report; but to the extent of its investigation and within the scope of the work delegated to it, Kilborn believes that the report is substantially correct.

Ontario Lignite
Indirect Liquefaction Study
Phase 1
Gasification Process Selection

Prepared for:
Ontario Energy Corporation

Kilborn Limited
August 1983

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Introduction

1.0 INTRODUCTION

This phase of the study is a further evaluation of the gasification processes which were assessed in Phase 0. The processes were determined in Phase 0 to be suitable for the gasification of Ontario Lignites to produce a synthesis gas for downstream processing to methanol. The objective of the Phase 1 program was to select a maximum of two gasification/conversion processes most financially attractive for the conversion of Ontario Lignite to methanol. The selected process(es) will be the subject of more detailed study through lignite testing, process design and cost estimating during Phases 2 and 3 of the study.

Phase 0 was completed in March 1983 and assessed the following gasification technologies relevant to their potential to gasify Ontario lignites.

- Koppers-Totzek Entrained Bed Process
- Lurgi Fixed Bed Process
- Westinghouse Fluidized Bed Process
- High Temperature Winkler Fluidized Bed Process

Each technology was reviewed with respect to its ability to gasify Ontario lignites placing emphasis on the following parameters and considerations:

a) Predicted behaviour of Ontario lignite in the gasifiers considering proximate, ultimate, ash and petrographic analyses of the lignite. These analyses were provided to Kilborn by Ontario Energy Corporation. Particular emphasis was placed on moisture content, ash content and ash fusion temperature characteristics of the Ontario lignite.

b) Expected syngas composition and downstream processing requirements typical to each gasification process.

c) Extent of commercial and/or pilot plant experience, particularly with lignites displaying similar properties to Ontario lignites.

d) Process and operating performance assessment with emphasis on the following aspects:

  - Lignite preparation for efficient processing,
  - Expected yield, carbon conversion and thermal efficiency,
  - Start-up, shut-down and operating flexibility of gasifiers,
  - Mechanical and materials performance.

e) Environmental requirements in order to meet the applicable Federal and Provincial guidelines.

f) Contact with each technology developer to obtain their initial prediction of gasification potential based on the analyses of the Ontario lignite.

All four of the technologies were recommended to be carried forward to Phase 1 for the economic assessment. Each process was considered suitable to gasify Ontario lignites based on the technology review and the predictions made by the developers. It was recommended to place particular emphasis in Phase 1 on the following aspects:

- Lignite predrying requirements and the possible effects on fines generation, particularly with respect to the Lurgi process. Lurgi (the process developer) indicated that, due to the high moisture content of the Ontario lignites, severe decrepetation of lignite may occur during predrying, causing excessive fines generation which cannot be handled efficiently in the gasifiers.

- Synthesis gas processing options to produce an optimum H_2/CO ratio for methanol synthesis.
- Differences in effluent treatment facilities to meet environmental regulations.

This report presents a financial comparison of the four gasification technologies, including the downstream processing requirements to produce fuel grade methanol as the end-product. The assessment encompasses all processing steps in the conversion of lignite to methanol fuel including:

- lignite preparation (crushing and drying),
- lignite gasification,
- ash/slag handling,
- synthesis gas cooling,
- liquids separation,
- shift conversion,
- gas treating (H_2S and CO_2 removal),
- methanol synthesis,
- methanol purge gas treating,
- methanol purification,
- gasification liquid effluent treatment, tar oil recovery and ammonia recovery,
- oxygen plant,
- H_2S and tail gas treating to produce sulphur,
- utilities,
- offsites.

In order to perform the financial comparisons, preliminary capital and operating cost estimates have been prepared in first quarter 1983 Canadian dollars for a Northern Ontario location. Cost estimates were prepared for each technology for a base case and various options for lignite-to-methanol plants. The base case and options were discussed in detail in Section 3.0 of the Phase 0 report. The base case scenarios for this report are identical to those presented in the Phase 0 report.

b) Expected syngas composition and downstream processing requirements typical to each gasification process.

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d) Process and operating performance assessment with emphasis on the following aspects:

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- methanol purge gas treating,
- methanol purification,
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However, there were slight modifications made to the options as it became apparent during the process design and subsequent costing that some optimization was required in order to achieve the most financially attractive result. The description of the base cases and options for each gasification technology is presented in Section 3.0.

For the financial comparison to be consistent, a 2000 t/day pure methanol plant was initially selected, which corresponds to 2010 t/day of fuel grade methanol. The preliminary design for the base cases has been performed compatible with this capacity. However, for the options, particularly the co-processing scenario of lignite and natural gas being converted to a syngas for methanol synthesis, capacities different than 2010 t/day resulted. The different capacities have been equalized by computing a capital cost per tonne of methanol for each technology under consideration. The process design basis is presented in Section 3.0 of this report.

In Section 4.0 of the report, the capital and operating cost estimates are presented for the base case and process options for each technology. From these estimates, production and product costs have been computed per tonne of methanol product. Technical and financial assessments have been performed based on the data presented in the report and recommendations made to pursue specific test programs with process developers in order to provide pertinent process and financial data for Phase 3 of the study.

Conclusions and Recommendations

2.0 CONCLUSIONS AND RECOMMENDATIONS

The conclusions and recommendations resulting from the Phase 1 assessment of the four gasification technologies are highlighted as follows:

2.1 CONCLUSIONS

2.1.1 Process Technologies

.1 Financial Assessment

A summary of the base case capital and operating costs, and total product cost is presented for the four technologies in Table 2-1.

The main criterion used in assessing the four technologies is the total product cost of methanol. This cost is determined by combining the operating cost per tonne of methanol with a capital cost per tonne of methanol. The capital cost per tonne was determined for three capital recovery factors; 10, 15 and 20 percent. Ten percent is representative of a public utility investment criterion while 20 percent is representative of a private investment criterion.

The relative ranking of the four technologies based on product cost is as follows:

- Westinghouse
- High Temperature Winkler
- Lurgi
- Koppers-Totzek

TABLE 2-1
BASE CASE CAPITAL AND OPERATING COSTS

	KOPPERS-TOTZEK	LURGI	WESTINGHOUSE	HIGH TEMPERATURE WINKLER
Methanol Capacity, t/y	643 200	643 200	643 200	643 200
Capital Cost, $ million	691.2	681.1	639.1	697.6
Capital Cost per annual tonne methanol, $/tonne	1074.6	1058.9	993.6	1084.6
Operating Cost, $/tonne	119.39	117.69	102.92	109.13
By-Product Credit, $/tonne	2.34	6.62	3.22	4.22
Total Production Cost, $/tonne (Operating Cost minus By-Product Credit)	116.75	111.07	99.70	104.91
Total Product Cost, $/tonne [Production cost +(Capital cost/t x percent capital recovery)]				
at 10% capital recovery	224.51	216.96	199.06	213.37
at 15% capital recovery	278.24	269.91	248.74	267.60
at 20% capital recovery	331.97	322.85	298.42	321.83

Capital Cost

While there is no significant difference between Lurgi, Koppers-Totzek and High Temperature Winkler capital costs, the capital cost of the Westinghouse process was found to be marginally lower than the others.

The main difference between the costs of the Westinghouse and H.T. Winkler technologies is in the installed cost of the gasification facilities provided by the process developers. Since both technologies are fluidized bed processes, one would assume the capital costs to be competitive. Indeed, the Westinghouse gasifiers should have a higher capital cost as they operate at a higher pressure than H.T. Winkler gasifiers. This apparent discrepancy was not resolved within the time frame of the Phase I report.

The difference in capital cost between the Westinghouse and Lurgi processes is due to the additional downstream processing, effluent treatment and utility requirements required in the Lurgi process. The difference in capital cost between Westinghouse and Koppers-Totzek is due to the extensive syngas compression facilities needed for the Koppers-Totzek process, entailing large steam turbine driven compressors.

Operating Cost

There is no significant difference between the production costs of the Westinghouse and H.T. Winkler processes. This is to be expected as the process configurations of these technologies are similar. The higher production costs for Lurgi and Koppers-Totzek are due to the higher process steam consumption for Lurgi and the high turbine steam requirements for Koppers-Totzek.

A zero by-product credit was assessed to the lignite fines in the Lurgi process. For purposes of this study it was assumed that lignite fines in excess of those used for steam generation would be sold at zero value (F.O.B. plant) for thermal power generation.

.2 Technical Assessment

As many of the technical advantages or disadvantages relevant to the four competing processes are reflected in the capital and operating costs, a detailed technical ranking has not been performed. However, there are criteria that are not reflected in the financial assessment, such as stage of development, experience with lignites and equipment performance, which are briefly discussed below.

Stage of Development

The Koppers-Totzek and Lurgi processes are commercially proven processes and hence meet the study criteria that the accepted technology must be commercially available by 1990. The Koppers-Totzek four headed gasifier has been in commercial use since the late 1970's while Lurgi gasifiers have operated commercially since the mid 1950's. The Lurgi Mark IV units have been operating at SASOL, South Africa, since 1958 where over 80 units have been installed.

The Westinghouse and High Temperature Winkler gasification technologies are not yet commercially available. However, as emerging technologies these processes have been demonstrated in bench scale and/or pilot plant facilities. The Westinghouse technology was to be demonstrated commercially by late 1986 in a demonstration project in SASOL. The demonstration unit would be capable of processing 1090 t/d of sub-butuminous coal. However, in July 1983 this project was cancelled due to the current energy supply situation. This is a serious setback to the development of the Westinghouse technology as it appears that commercialization may not be achieved before 1990.

In fact, an announcement in the August 8, 1983 issue of Chemical Engineering indicated that Westinghouse is seeking buyers for its Synthetic Fuels Division as falling oil prices and escalating commercialization costs have made the technology unattractive. The High Temperature Winkler process has been tested extensively in pilot plant operation and the first commercial size demonstration plant is due to begin operation in Wesseling, West Germany in 1984.

Experience with Lignites

The Lurgi process has been tested extensively with lignite with a full-scale test carried out in 1974 on North Dakota Lignite at SASOL's facilities. This test confirmed that a high proportion of the lignite feed, 25-35 percent, is generated as fines in the preparation process and hence is unsuitable for gasification. This is not a serious drawback if efficient use can be made of the fines. However in the Onakawana location this is a serious concern as an excess of fines exists after consumption for steam and power generation, and hence a byproduct credit cannot be assessed to the fines.

The Koppers-Totzek process requires a pulverized lignite feed of 8 percent moisture, maximum. Drying high moisture Onakawana lignite to such a low level involves a high energy consumption and significant safety hazard owing to the risk of spontaneous combustion and/or explosion.

Westinghouse has tested Texas and North Dakota lignites in its Waltz Mill, Pennsylvania pilot plant facility. Westinghouse has reported that these lignites can be efficiently gasified to yield a syngas suitable for downstream processing applications. Rheinbraun, the High Temperature Winkler process developer, has tested North Dakota and German lignite in its pilot plant facility, and the demonstration plant in Wesseling will use German brown coal, a lignite, as the feedstock.

Material and Equipment Performance

Since the Koppers-Totzek and Lurgi gasification processes are commercialized and the Westinghouse and High Temperature Winkler process are in the development stage, it is difficult to rate each process with respect to material and equipment performance due to the lack of data on the developing technologies. Consequently, only the key areas of concern can be highlighted.

The Koppers-Totzek process operates in the slagging mode at a much higher temperature than the competing gasifiers (1500-1600°C versus 1100°C for H.T. Winkler). The integrity of the refractory lining in contact with the high sodium and potassium Onakawana lignite slag is a concern as severe refractory problems have been encountered with certain slags. If frequent refractory failure results, an increase in plant downtime is inevitable unless gasifier sparing philosophy is revised.

2.1.2 Natural Gas Co-Processing

Of the four gasification processes reviewed in this phase of the study only the Lurgi process had an H_2:CO ratio in the raw syngas suitable for methanol synthesis. The other gasification processes were deficient in hydrogen and hence shift conversion of the syngas was necessary to increase the hydrogen content. An alternative method of adjusting the H_2:CO ratio is by co-processing natural gas with lignite. In general, co-processing with natural gas results in higher overall thermal efficiencies. Two options where investigated in Phase 1: Option 1, steam reforming of natural gas to a hydrogen-rich syngas for blending with gasifier raw syngas and Option 2, steam reforming of natural gas followed by Pressure Swing Adsorption to supply high purity hydrogen for blending with gasifier raw syngas.

Capital, operating and product costs for the options considered are presented in Table 2-2. The capital costs per product tonne for these plants are lower than for the base cases as costly gasification sections are replaced by lower cost steam reforming facilities.

Operating costs are much higher as low cost lignite is replaced by high cost natural gas as the major feedstock. The combined cost, as represented by the total product cost is lowest for the H.T. Winkler process for the following reasons:

- Less syngas from natural gas reforming is required to adjust the combined gas H_2:CO ratio for methanol synthesis, with subsequent reductions in capital and operating costs for natural gas processing facilities.

- The H.T. Winkler gasifier syngas output, when supplemented with syngas from natural gas processing facilities, results in a methanol plant of 2781 t/d. Hence, on a tonne/day basis the total product cost is favoured by economies of scale. This product cost tends to be lower since it was assumed for the study that 3000 t/day methanol can be synthesized in a single train unit. To date, 2500 t/d is the largest commercial single train methanol plant, but 3600 t/d methanol capacity has been proposed for single train operation in the conceptual design of a Lurgi methanol synthesis process[1].

As mentioned previously, both options provide a lower methanol product cost than conversion of lignite only. However, the cost of natural gas for these options is about 30-40 percent of the total product cost.

2.1.3 Electrolytic Hydrogen

Co-processing of lignite and electrolytic hydrogen was determined to have potential as an alternative to lignite only conversion. Further, it was determined that purchasing power for hydrogen electrolysis is preferable to onsite co-generation of power.

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[1] "Conceptual Design of a Coal-to-Methanol-to Gasoline Commercial Plant" - Badger Plants Report for United States Department of Energy, March 1979.

Similar to natural gas co-processing, electrolytic hydrogen-lignite co-processing is an alternative method of adjusting the H_2:CO ratio in the Koppers-Totzek, Westinghouse and H.T. Winkler processes. Two options were considered: Option 3, electrolysis of water using purchased power to produce oxygen for gasification and hydrogen for blending with the gasifier raw syngas and Option 4, electrolysis of water using onsite co-generated power to produce oxygen for gasification and hydrogen for blending with the gasifier raw syngas, for the Westinghouse process.

Capital, operating and product costs for these options are presented in Table 2-3. The capital costs per product tonne for Option 3 are lower than the base case costs since a portion of the shift conversion, air separation and gas treating facilities are replaced by lower cost electrolysis facilities. Operating costs are much higher for Option 3 than for the base case, due mainly to the purchased electricity cost to supply power to the electrolysis plant. Conversely, Option 4 has a higher capital cost per product tonne due to the addition of power generation facilities and lower operating costs since power is generated onsite.

Due to the very high capital cost of power generation facilities the total product cost for Option 4 is significantly higher than Option 3 in which power is purchased from the provincial grid. Hence onsite power generation is not an attractive alternative to purchasing power.

In general, the total product cost associated with hydrogen electrolysis is significantly higher than the lignite only base cases. There is, however, potential for a reduction in the cost of power based on the very high demand (around 250 MW) required for an electrolysis plant. The cost of power used in Phase 1 (2.71c/kWh) was provided by Ontario Hydro. However, there are incentive programs underway in Northern Ontario which may provide favourable utility rates in order to attract industrial development. It is feasible that a rate of 1.8¢ to 2.0¢/kWh may be negotiated for such demand in Northern Ontario.

TABLE 2-2

NATURAL GAS CO-PROCESSING CAPITAL & OPERATING COSTS

	KOPPERS-TOTZEK		WESTINGHOUSE		H.T. WINKLER	
	Option 1	Option 2	Option 1	Option 2	Option 1	Option 2
Methanol Capacity, t/y	847 360	802 960	712 640	679 650	889 600	898 690
Capital Cost, $ Million	533.0	622.9	451.9	536.4	550.2	672.5
Capital Cost per Annual tonne Methanol, $/tonne	629.0	775.7	634.1	789.2	618.5	748.3
Operating Cost, $/tonne	162.39	149.44	157.71	143.59	155.43	139.21
By-Product Credit, $/tonne	.42	.91	.73	1.54	1.04	1.99
Total Production Cost, $/tonne (Operating cost minus By-Product Credit)	161.97	148.53	156.98	142.05	154.39	131.09
Total Product Cost, $/tonne [Production cost +(Capital cost/t x percent Capital recovery)]						
at 10% capital recovery	224.87	226.10	220.39	220.97	216.24	212.05
at 15% capital recovery	256.32	264.88	252.09	260.43	247.16	249.47
at 20% capital recovery	287.77	303.67	283.80	299.89	278.09	286.88

In addition, electrolytic hydrogen cell technology is still developing and substantial improvements and capital cost reductions are foreseen in the future. The capital costs of electrolysis plants used in this study are conservative and are based on rough estimates provided by Electrolyser Corporation. Electrolyser Corporation provided an installed cost of $300/kW which was based on a scaled cost of an experimental electrolysis plant in Varennes, Quebec. It is possible that, with technological advances in the unipolar cell technology, the cost of a large electrolysis plant could approach $200/kW by 1985.

These capital and operating cost reductions could potentially make this case attractive. With favourable power rates and a reduction in the capital cost of electrolysis plants to $200/kW, it is estimated that methanol could be produced for $30-35/tonne lower than indicated, or for about the same cost as the lignite only base cases.

2.1.4 Partial Oxidation

In gasifier process configurations where the methanol synthesis purge gas contains substantial quantities of methane, steam reforming of the purge gas to yield a hydrogen-rich syngas is preferable to partial oxidation of the purge gas.

In the Lurgi and Westinghouse base cases a portion of the purge gas is steam reformed to a hydrogen-rich syngas for recycle to the methanol synthesis unit. An alternative method of processing the purge gas is partial oxidation. To test this alternative, a special case designated Option A was prepared for the Lurgi process configuration.

Capital, operating and product costs for the Lurgi base case and Option A is presented in Table 2-4. As illustrated in Table 2-4, both the capital and operating costs for the partial oxidation option are higher than the steam reforming base case. As a result, steam reforming of purge gas is preferable to partial oxidation.

TABLE 2-3

ELECTROLYTIC HYDROGEN CO-PROCESSING CAPITAL & OPERATING COSTS

	KOPPERS-TOTZEK	WESTINGHOUSE		H.T. WINKLER
	Option 3	Option 3	Option 4	Option 3
Methanol Capacity, t/y	629 130	629 130	629 130	629 130
Capital Cost, $ million	655.3	589.0	915.3	629.2
Capital Cost per Annual tonne Methanol, $/t	1041.6	936.2	1454.9	1000.0
Operating Cost, $/tonne	161.80	148.77	119.52	167.52
By-Product Credit, $/tonne	1.74	2.18	2.18	2.85
Total Production Cost, $/tonne (Operating cost minus By-Product Credit)	160.06	146.59	117.34	164.67
Total Product Cost, $/tonne [Production cost + (Capital cost/t x percent capital recovery)]				
at 10% capital recovery	264.22	240.21	262.89	264.67
at 15% capital recovery	316.30	287.02	335.58	314.67
at 20% capital recovery	368.38	333.83	408.32	364.67

TABLE 2-4

PARTIAL OXIDATION/STEAM REFORMING CAPITAL AND OPERATING COSTS

	LURGI	
	Base Case Steam Reforming	Option A Partial Oxidation
Methanol Capacity, t/y	643 200	643 200
Capital Cost, $ Million	681.1	687.5
Capital Cost per Annual tonne Methanol, $/tonne	1058.9	1068.9
Operating Cost, $/tonne	117.69	120.17
By-Product credit, $/tonne	6.62	6.62
Total Production Cost, $/tonne (Operating cost minus By-Product Credit)	111.07	113.55
Total Product Cost, $/tonne [Production cost + (Capital cost/t x percent capital recovery)]		
at 10% capital recovery	216.96	220.44
at 15% capital recovery	269.91	273.89
at 20% capital recovery	322.85	327.33

2.2 RECOMMENDATIONS

2.2.1 Selected Process

.1 It is recommended that the High Temperature Winkler process be evaluated in the Phase 2 test program. This recommendation has been made for the following reasons:

- The High Temperature Winkler process appears to be the most financially and technically attractive technology which meets the study requirement of being commercially proven by 1990.

- The Westinghouse technology is the most financially attractive process, but the current announcement concerning the cancellation of the commercial demonstration project casts doubt on the future of the technology development. It is now doubtful that the Westinghouse technology will be commercially proven by 1990.

- The financial assessment shows that the Westinghouse process has the lowest product cost. This is because the Westinghouse gasification unit cost is much lower than the High Temperature Winkler ($85.8 million versus $132 million for base case). As both processes are fluidized bed technologies, one would expect the costs to be competitive. If the processes were to be assessed utilizing the same installed costs for the gasification systems, the H.T. Winkler process would be the most financially attractive overall.

- The financial and technical assessments made for all process design options indicate that the fluidized bed technologies, Westinghouse and High Temperature Winkler, are the preferred technologies. The product costs per tonne of methanol and overall process thermal efficiencies are all superior to the Lurgi and Koppers-Totzek technologies for the base case and are, for the most part, superior for the optional cases.

- The most financially attractive, commercially proven process is the Lurgi process. However, there was a serious concern expressed by the process developer that handling and drying the Onakawana lignite would result in the generation of excessive fines which cannot be processed in the Lurgi gasifier. A fines generation level of 25 percent was assumed for this study. This may be optimistic as 30-35% fines are assumed to be generated from North Dakota lignite feed to Lurgi gasifiers for the Great Plains Gasification Project. The North Dakota lignite is submitted to size reduction only, as drying is not required. Fines disposal is not a problem for the Great Plains Project as the fines are fed directly to an adjacent thermal power plant. A by-product credit of zero value has been assessed to the fines in this study due to the uncertainty of a market to a generating station. Simulated size reduction and drying tests would be required to confirm the extent of fines generation before testing with the process developer could be considered. As it is expected that tests will confirm that fines generation will exceed the 25 percent assumed for the study, it is recommended that the Lurgi process should be eliminated from further evaluation in this study.

- At a 20 percent capital recovery factor, the commercially proven Koppers-Totzek process was consistently the highest cost process of the four technologies evaluated. This is mainly due to the high cost of the syngas compression facilities. This gasification process operates at a severe operating temperature (about 1600°C) and there is a concern that the high sodium and potassium levels in Onakawana lignite ash could cause premature refractory failure in Koppers-Totzek process gasifiers. Consequently it is recommended that the Koppers-Totzek not be considered for the Phase 2 test program.

.2 It is recommended that the preferred processing option (lignite only, lignite/natural gas co-processing or lignite/electrolytic hydrogen) be evaluated after the test program has been completed at Rheinbraun's facilities and more definitive data on syngas composition has been received.

2.2.2 Test Program

Two drums of Onakawana lignite, identified as Samples 1 and 13 comprise the testing material. The weight of lignite available is estimated at 310 kg at a predicted 30-35% moisture content, equivalent to 240-250 kg at 15% moisture content. The test program recommended by Rheinbraun Consulting GmbH consists of three stages as follows:

Stage I

a) Physical/chemical examination of the lignite to determine ash melting behaviour, reactivity, pore structure, ignition temperature and particle distribution.

b) Cold fluidized bed tests to determine fluidizing velocity and flow behaviour in the fluidized bed.

Lignite required: 50 kg

Stage II

In addition to the Stage I test, a hot gasification test to determine the general suitability of the lignite in a fluidized bed, approximate carbon conversion and product gas composition.

Lignite required: 250 kg

Stage III

In addition to Stages I and II this involves gasification tests in a 35 kg/hr pilot plant to confirm the determinations in Stage II above and to provide preliminary energy and mass balances.

Lignite required: 3700 kg

The costs for these tests are as follows:

Stage I	34 170 DM ($ 15 800)
Stage II (Includes Stage I)	65 500 DM ($ 30 400)
Stage III (Includes Stages I & II)	286 000 DM ($132 600)

1 DM = .4635 $ Canadian.

As there is insufficient lignite to fulfill the requirements of Stage III and as the cost exceeds the budget allocated for Phase 2 of the study, it is recommended that the test program up to and including Stage II be completed.

The tests would be performed at Aachen University and at the Rheinbraun pilot plant facilities in Frechen (near Cologne), West Germany.

The expected duration of testing and issue of a report is 10-12 weeks.

Technical Description

3.0 TECHNICAL DESCRIPTION

3.1 PROCESS DESIGN CRITERIA

3.1.1 General

This section describes the four gasification processes which were selected for further study in Phase 0 of this study. For each gasification process several downstream processing options were considered including lignite only conversion, lignite plus natural gas co-processing and lignite plus electrolytic H_2 conversion. Table 3-1 lists the options considered. Block Flow Diagrams, overall material balances and utility requirements are presented for each process configuration.

The lignite analysis used as the basis for this phase of the study is an arithmetic average of the analyses presented in Phase 0 of the study for Onakawana LTM-03 Sample #1 and Sample #2. In addition, a moisture content of 45% was assumed. The proximate and ultimate analyses of the lignite are provided in Table 3-2.

The base case process configurations for each gasifier were sized to convert lignite only to 2010 tonne per day of fuel grade methanol (2000 t/d of pure methanol). This capacity is considered representative of a conventional world scale natural gas to methanol plant.

3.1.2 Downstream Process Technology

In all four gasifier processes lignite is converted with oxygen (and steam for Lurgi, Westinghouse and H.T. Winkler processes) to a raw syngas. The raw syngas, depending upon its H_2:CO ratio, is shifted to a correct H_2:CO ratio and fed to an acid gas removal unit. A Lurgi Rectisol acid gas removal unit is used as a representative method of selectively removing CO_2 and H_2S from the raw syngas. Both Norton and Union Carbide offer competing acid gas removal processes, Selexol and Benfield, respectively. Both of these alternatives offer advantages; however, selection of the process best suited to this project should be based on a technical and financial comparison of the competing processes at a later date.

TABLE 3-1
PROCESS OPTIONS

GASIFIER	OPTION	DESCRIPTION	FUEL GRADE METHANOL CAPACITY t/d
Koppers-Totzek	Base Case	Lignite Only	2010
	Option 1	Lignite-Natural Gas Syngas	2648
	Option 2	Lignite-Natural Gas H_2	2533
	Option 3	Lignite-Electrolytic H_2	2010
Lurgi	Base Case	Lignite Only - Purge Gas SMR	2010
	Option A	Lignite Only - Purge Gas POX	2010
Westinghouse	Base Case	Lignite Only	2010
	Option 1	Lignite-Natural Gas Syngas	2227
	Option 2	Lignite-Natural Gas H_2	2144
	Option 3	Lignite-Electrolytic H_2	2010
	Option 4	Lignite-Electrolytic H_2 + Cogeneration	2010
High Temperature Winkler	Base Case	Lignite Only	2010
	Option 1	Lignite-Natural Gas Syngas	2780
	Option 2	Lignite-Natural Gas H_2	2835
	Option 3	Lignite-Electrolytic H_2	2010

TABLE 3-2
ONAKAWANA LIGNITE ANALYSIS

PROXIMATE ANALYSIS, wt%	
Moisture	45.0
Ash, db	18.81
Volatile Matter, db	40.49
Fixed Carbon, db	40.70
Heating Value, kJ/kg, db	22348
ULTIMATE ANALYSIS, wt% db	
Carbon	59.30
Hydrogen	3.81
Oxygen	15.97
Nitrogen	.70
Sulphur	1.41
Ash	18.81

Concentrated H_2S from the acid gas removal unit is converted to elemental sulphur in a conventional three stage Claus sulphur plant. A SCOT tail gas unit was selected as typical of competing tail gas units to reduce sulphur emissions from the Claus plant.

Clean syngas is converted to methanol in a methanol synthesis unit based on Lurgi low pressure technology. This process is considered typical of current methanol technology and hence suitable for this phase of the study. Both Imperial Chemical Industries and Mitsubishi Gas Chemical offer competing methanol synthesis processes. Final selection of the methanol process best suited for this project should be based on a technical and financial comparison of the processes at a later date.

A ten percent excess of hydrogen is provided in the methanol feed gas to maintain high conversion rates in the unit. Accordingly, process developers data were adjusted as necessary.

3.1.3 Base Case Considerations

While all four gasifiers convert lignite to a syngas suitable for methanol production, the Lurgi and Westinghouse gasifiers produce appreciable quantities of methane. To improve the energy efficiency of the complexes, methane-rich purge gas from the methanol synthesis unit is split and a portion reformed with steam to a hydrogen-rich syngas. The syngas is recycled to the methanol synthesis feed. Unconverted purge gas is utilized as fuel gas in the complex.

In addition to high methane content in the raw syngas, because of its low gasification temperature, the Lurgi gasifier produces tars, tar oils, phenols and naphtha. These components are recovered in a relatively complex recovery system and combusted for steam generation. Although these compounds could be stored and marketed, the remote location of the complex and subsequent high transportation cost to markets preclude this option.

The Koppers-Totzek and High Temperature Winkler gasifiers produce a syngas rich in hydrogen, carbon monoxide and carbon dioxide. Hence special downstream treatment, such as steam reforming of the methanol synthesis purge gas, is unnecessary.

3.1.4 Natural Gas Co-processing

Of the four gasifiers considered in this evaluation, only Lurgi produced a satisfactory H_2:CO ratio in the raw syngas for methanol production. Syngas from the Westinghouse, Koppers-Totzek and H.T. Winkler gasifiers must be shifted to increase the hydrogen content in the syngas.

Another method of increasing hydrogen content in the syngas is to add additional hydrogen from a natural gas source. For the above mentioned gasifiers two optional process configurations were considered:

Option 1: Steam reforming of natural gas to a hydrogen rich syngas. This syngas was blended with gasifier raw syngas in an appropriate ratio to obtain the desired H_2:CO ratio for methanol synthesis.

Option 2: Steam reformed natural gas is shifted to maximum hydrogen content and fed to a Pressure Swing Adsorption (PSA) unit. High purity hydrogen from the PSA unit is blended with gasifier syngas to obtain the desired H_2:CO ratio for methanol synthesis.

Both co-processing options produce an excess of methanol synthesis feed gas. Thus it was decided to reduce the number of gasifiers in the optional cases rather than the throughput per gasifier. In this way methanol production was allowed to vary between 2000 and 3000 tonne per day depending upon the gasifier throughput and optional case considered.

3.1.5 Electrolytic Hydrogen

An alternate method of increasing the hydrogen content in the syngas of the Westinghouse, Koppers-Totzek and H.T. Winkler gasifiers is to add electrolytic hydrogen. For these gasifiers two additional process configurations were considered:

Option 3: Electrolysis of water to yield high purity hydrogen and oxygen. Oxygen is sent to the gasifier while the hydrogen stream is blended with gasifier syngas to obtain the desired H_2:CO ratio for methanol synthesis. Power for electrolysis is purchased.

Option 4: Electrolysis of water similar to Option 3 but power is produced onsite by co-generation using lignite as fuel, evaluated for the Westinghouse gasifier only.

Similar to the co-processing options, an excess of methanol synthesis feed gas is produced by these options. Thus it was decided to reduce the number of gasifiers in these cases.

3.1.6 Lurgi Partial Oxidation

In the Lurgi base case configuration, methane rich purge gas from the methanol synthesis unit is split and a portion steam reformed to a hydrogen rich syngas for recycle to the methanol synthesis unit. An optional process configuration was considered for this gasifier:

Option A: A portion of the methanol synthesis purge gas is fed to a partial oxidation (POX) unit. Partial combustion of the purge gas with oxygen yields a hydrogen rich syngas suitable for recycle to the methanol synthesis unit.

3.1.7 Utilities

While every effort has been made to minimize energy consumption in the processes reviewed, some of the options required separate steam generation facilities in addition to the waste heat recovery systems within the complex. It was decided that lignite was the most suitable fuel for offsite steam generation since it would only require an incremental increase in the overall lignite demand. Purge gas from the methanol synthesis and purification units is also used when available for steam generation. For the natural gas co-processing option the fuel to the reformer is provided by recycled purge gas supplemented if required by imported natural gas.

For all process options except the Westinghouse electrolytic H_2-cogeneration case, it was assumed that power would be purchased from the provincial grid. In addition, raw water was assumed to be available in abundant quantities. There is a concern that water supplies could be insufficient in the summer. However, verifying and assessing this concern is beyond the scope of this phase of the study.

3.1.8 Operating Factors

In general, it is assumed from the operating experience of large scale coal gasification complexes that a gasification complex with suitable spare gasifiers should be able to operate 320 days per year. The remaining 45 days represent scheduled and unscheduled down time of the entire complex.

In the optional process configurations for Westinghouse, Koppers-Totzek and H.T. Winkler gasifiers the number of operating gasifier trains, as mentioned previously, was reduced. An assessment of the percentage of design methanol capacity that could be obtained should one operating gasifier be shut down is illustrated in Table 3-3. It was decided that gasifier options that would operate at greater than 65 percent of design capacity did not require a spare gasifier; gasifier options that would operate below 65 percent were assigned one spare gasifier.

Gasifier options that were assigned a spare gasifier were assumed to have an operating factor of 320 days. All other gasifier options were assigned an operating factor based on a penalty for reduced throughputs should one gasifier be shutdown. Operating factors are provided in Table 3-3 and illustrated in the following example:

Westinghouse Option 3 - Electrolytic H_2

- Assume	: General Plant Shutdown	45 days
	: 100% Design Capacity	290 days
	: 65% Design Capacity (1 gasifier shutdown)	30 days
		365 days

- Adjusted Operating Factor (290 x 1.0) + (30 x 0.65) =	310 days

TABLE 3-3
SUMMARY OF OPERATING FACTORS

Gasifier	Process Option	Daily MeOH Capacity, t/d	Number of Operating Gasifiers	% of Design Capacity[1]	Adjusted Operating Factor (days)	Additional Spare Gasifiers
Koppers-Totzek	Base Case	2010	4	75	320	1 Spare
	Option 1	2648	1	52.5	320	1 Spare
	Option 2	2533	2	91.0	317	No Spare
	Option 3	2010	3	75.0[2]	313	No Spare
Lurgi	Base Case	2010	5	80	320	1 Spare
	Option A	2010	5	80	320	1 Spare
Westinghouse	Base Case	2010	4	96	320	No Spare
	Option 1	2227	1	50.5	320	1 Spare
	Option 2	2144	2	88.5	317	No Spare
	Option 3	2010	2	65.0[2]	313	No Spare
	Option 4	2010	2	65.0[2]	313	No Spare
H.T. Winkler	Base Case	2010	3	67	320	1 Spare
	Option 1	2780	1	48.0	320	1 Spare
	Option 2	2835	2	83.0	317	No Spare
	Option 3	2010	2	67.5[2]	313	No Spare

Notes

[1] % of design capacity should one operating gasifier be shut down and spare gasifier (where applicable) not started up.

[2] In the event of one gasifier being shut down, the shift conversion unit can be bypassed to optimize the H_2/CO ratio

3.2 KOPPERS TOTZEK GASIFICATION TECHNOLOGY

3.2.1 Introduction

The Koppers-Totzek gasification technology has been in commercial use since 1952. The Koppers-Totzek gasifiers can process a wide variety of fuels including bituminous and sub-bituminous coals, lignite, peat, heavy petroleum residue and petroleum coke; the most common end product being ammonia. Lignite is converted to low BTU gas and ammonia in a plant in Ptolemais, Greece. The largest scale application of Koppers-Totzek gasifiers for methanol production is at AECI, Modderfontein, South Africa where sub-bituminous coal is used as the gasifier feedstock.

The Phase 0 report reviewed the operating and maintenance characteristics of the Koppers-Totzek process in detail; hence, only a summary of the advantages and disadvantages will be presented in this report.

Advantages

(i) the process can handle a wide variation in coal properties

(ii) The process operates at a sufficiently high temperature (gas temperature of 1400-1600°C), to prevent tar, phenol or condensible hydrocarbon production

(iii) the equipment is simple and capacities are flexible: full capacity can be attained within 30-60 minutes from start-up.

Disadvantages

(i) The Koppers-Totzek gasification system operates only slightly above atmospheric pressure. Consequently, there is a very high cost penalty for compression of the syngas to meet downstream processing requirements (5500 kPa minimum for methanol production).

(ii) Gasifier capacities are lower than comparable gasification processes. This drawback has been tempered with the introduction of the four headed gasifier which is designed to process about 800 t/d of coal per gasifier. This compares favourably with the Lurgi MARK IV gasifier but is below the capacities of the Westinghouse and High-Temperature Winkler gasifiers.

(iii) Due to the high temperature operation of the KT gasifier, sodium and potassium present in the slag tends to attack the refractory lining in the gasifier and foul the waste heat boiler immediately downstream of the gasifier.

(iv) In order to operate at a reasonable thermal efficiency the process requires a very low moisture coal feed which, for lignite, should be less than 8 percent. Consequently, there is a high investment cost for crushing and drying operations particularly for Onakawana lignite which has a high moisture content (45-50 percent). Also, the drying of lignite to such a low moisture content incurs a significant safety hazard, such as spontaneous combustion and explosion risks.

The following technical review primarily describes the facilities for the base case process. Only areas which are unique to the optional cases are described in detail, while areas with facilities similar to the base case are not described.

Block Flow Diagrams for the base case and options are included as follows:

Base Case

Drawing 100-43-01 describes a lignite to methanol plant with 2010 t/d methanol production. All the carbon requirements are provided by the lignite, and the H_2/CO ratio is adjusted by CO shift conversion.

Option 1

Drawing 100-43-02 describes a co-processing scenario of lignite and natural gas. Natural gas is delivered to the plant via a dedicated new pipeline from the Trans-Canada Pipeline. The natural gas is steam reformed to produce a hydrogen-rich syngas which is combined with the gasifier produced syngas for methanol synthesis. This option entails reducing the number of gasifiers from four to one and the total production of fuel grade methanol is 2648 t/d.

Option 2

Drawing 100-43-03 is a modification of Option 1. The syngas produced by natural gas reforming is further processed to obtain hydrogen in a pressure swing adsorption (PSA) unit. This hydrogen is added to the gasifier produced syngas for methanol synthesis. All the carbon requirements for methanol are provided by the lignite; thus two gasifiers are required. Methanol production with this option is 2533 t/d.

Option 3

Drawing 100-43-04 describes a lignite to methanol plant in which electrolytic hydrogen is used to supplement the hydrogen-deficient gasifier produced syngas. Capacity is identical to the base case at 2010 t/d of fuel grade methanol.

3.2.2 Facilities Description - Base Case

.1 Section 1100 Lignite Preparation Facilities

The lignite is prepared for gasification in two steps:

- Crushing (Area 1110)
- Drying (Area 1120)

Run-of-mine lignite, with a moisture content in the range of 45-50 percent, is stockpiled in a storage area for reclaim to the preparation plant.

The lignite is first fed to indirect rotary steam tube driers which utilize low pressure steam as a heating medium. The L.P. steam is generated as exhaust from the air separation plant air compression steam turbines. The moisture content of the lignite is reduced from 45 percent to about 22-25 percent in these dryers. The lignite is then fed to bowl mills which pulverize the lignite to 90 percent minus 74 micron (200 mesh). The lignite is dried to 8 percent moisture by circulating hot flue gases through the bowl mills. The flue gases for drying are provided by burning fuel gas produced in the methanol synthesis plant. Lignite is transported pneumatically with nitrogen to storage bins and stored in an inert atmosphere. The nitrogen is provided from the air separation plant.

.2 Section 1200 Gasification

The gasification section consists of the following areas:

Area 1210	-	Gasifier
1220	-	Waste Heat Boiler
1230	-	Washer Cooler
1240	-	Electrostatic Precipitators

Area 1210 Gasifier

Four operating gasifiers and one spare are required. Each gasifier is a four-headed steam-jacketed, refractory lined carbon steel shell capable of processing approximately 850-900 t/d of lignite (8% moisture), and employing burners spaced 90° apart.

The lignite is transported pneumatically from storage to feed bins which serve each head of the gasifiers. Lignite is metered from each feed bin by feed screws into a mixing zone where it is entrained in a stream of oxygen and fed to the gasifier.

In the gasifier the lignite is gasified virtually immediately (less than 1 second). The temperature of the resulting gas is about 1400-1600°C. Approximately 50% of the ash produced leaves the gasifier entrained with the product gas. The remaining ash leaves the gasifier through the bottom and is solidified by contact with water in a quench tank beneath the gasifier. Boiler feed water is passed through the gasifier jacket to generate low pressure steam (550 kPa).

Area 1220 Waste Heat Boiler (Raw Gas Cooling)

Product gas and ash particles exit the top of the gasifiers and are cooled in waste heat boilers which generate high pressure steam. Five waste heat boilers are required, one for each gasifier. The waste heat boiler consists of a radiant-surface exchanger followed by a fire tube exchanger. The fire tube exchanger is inclined from the exit of the radiant exchanger to the washer cooler.

Process conditions:

Gas flow (dry basis)	5279 t/d Nm^3/day
Inlet gas Temperature	1400-1600°C
Outlet Gas Temperature	175-260°C
Steam Generated	2150 t/d
Steam Pressure	10.8 MPa
Steam Temperature	480°C

Area 1230 Washer Cooler

Gas leaving the four operating waste heat boilers is combined and fed to a washer cooler. In the washer cooler the gas is cooled with circulating water to about 35°C and about 90 percent of the entrained particulates are removed. The solids are settled from the wash water, thickened to a slurry with about 15 percent solids and pumped to the settling ponds. Cool gas exits the washer cooler to the electrostatic precipitators.

Area 1240 Electrostatic Precipitators

Cool gas with a dust level of 4.6 mg/Nm³ is split into two streams and directed to two electrostatic precipitators to reduce the particulate loading to 0.23 mg/Nm³. Raw syngas leaves the electrostatic precipitators for compression and downstream processing.

.3 Section 1300 Syngas Compression, Shift Conversion and Acid Gas Removal

The raw gas exiting the gasification section has the following composition.

	mol % (mf basis)
H_2	28.4
CO	62.6
CO_2	7.50
CH_4	0.01
H_2S	0.69
N_2 + Ar	0.70
	100.00

Molecular Wt:	21.8
Pressure:	Slightly above atmospheric
Temperature:	35°C

Area 1310 Syngas Compression

The raw gas is compressed to about 3000 kPa in steam turbine driven compressors.

Area 1320 Shift Conversion

The syngas enters the shift reactor where the following reaction occurs:

$$CO + H_2O \longrightarrow CO_2 + H_2$$

The reactor contains a fixed bed with a sulphur resistant catalyst. Feed gas is preheated to about 270°C and exits the reactor at about 480°C due to the exothermic heat of the water/gas reaction. The water for the shift conversion reaction, approximately twice stiochiometric requirements, is supplied as 3.5 mPa medium pressure (MP) steam. Excess water after shift conversion is removed in knockout drums and the sour condensate treated for reuse.

The composition of the syngas exiting the shift conversion unit is as follows:

	mol %
H_2	47.05
CO	20.45
CO_2	31.36
CH_4	.01
H_2S	.48
N_2 + Ar	.65
	100.00

Area 1330 Acid Gas Removal

The acid gas removal unit employs the Lurgi Rectisol process which is designed to remove H_2S, COS, CO_2 and other impurities from the synthesis gas by means of methanol as a physical solvent. Because of the different solubilities of H_2S and CO_2 in methanol, these components may be recovered selectively. This permits concentration of the H_2S to sufficient levels for recovery of the sulphur in a conventional sulphur plant. After absorption of the gaseous impurities the methanol is regenerated and recycled for further absorption.

As the absorption capacity of methanol increases with decreasing temperature, temperatures in the order of -40°C are used. Regeneration of the methanol is accomplished by depressuring and distillation.

The gas feed composition to the methanol synthesis unit is:

	mol %
H_2	66.80
CO	29.00
CO_2	3.20
CH_4	0.01
N_2	0.99

.4 Section 1400 Methanol Synthesis and Purification

Area 1410 Methanol Synthesis

The methanol synthesis unit is based on the Lurgi low pressure process. The synthesis gas from the acid gas removal unit is first compressed to about 5500 kPa, preheated and fed to the methanol converter. In the reactor the synthesis gas is converted to methanol by the following reactions:

$$2H_2 + CO \qquad CH_3OH$$
$$3H_2 + CO_2 \qquad CH_3OH + H_2O$$

The Lurgi methanol converter is a tubular reactor with the synthesis catalyst in the tubes and a coolant in the shell to maintain isothermal conditions for methanol synthesis. The heat of reaction is removed by generating medium pressure steam in the converter shell. Synthesized methanol is condensed in downstream heat recovery systems and purge gases are recycled as fuel. The crude methanol containing 4.5 wt% water and hydrocarbons as impurities is transferred to in-plant storage for processing in the methanol purification area 1420.

Area 1420 Methanol Purification

In addition to water, a number of organic chemicals are synthesized and condensed with the crude methanol. These chemicals and the water are removed in the methanol purification unit. The unit consists of two columns, the topping column and the refining column. The topping column removes the light ends and the refining column removes the majority of the water. Water withdrawn from the base of the refining column is discharged to the water treatment plant. Fuel grade methanol is cooled and pumped to storage.

.5 Section 1500 Air Separation

Area 1510 Air Separation

An air separation plant with a design capacity of 2616 t/d of 99.5% purity oxygen is required. Oxygen requirements are approximately 80 kg per 100 kg of lignite gasified (moisture free basis). Air is first filtered and compressed in centrifugal compressors. Following compression the air is dried and passed through a scrubber to remove CO_2. Air is cooled to about -162°C by cross exchanging with cold nitrogen and oxygen product streams and finally expansion. Rectification columns separate the nitrogen and oxygen components due to their boiling point differential. Nitrogen has a boiling point of -195.5°C and oxygen -183°C. Oxygen is withdrawn from the rectification column as a gas and cools the incoming air before being transferred to the gasifiers. The bulk of the nitrogen is vented. The remainder is available at 700 kPa for the following uses:

- Transportation of pulverized lignite feed
- Miscellaneous blanketing and purging
- Instrument and plant "air".

.6 Section 1600 Sulphur Recovery and Sulphur Slating

The sulphur recovery system comprises the Claus Sulphur Unit and the SCOT Tail Gas Unit. The sulphur recovery system produces 47 t/d of molten sulphur and an environmentally acceptable tail gas which is vented to atmosphere.

Area 1610 Claus Sulphur Unit

The Claus sulphur recovery unit is licensed by the Amoco Production Company. The hydrogen sulphide-rich gas from the acid gas removal unit flows to the Claus unit thermal reactor where the hydrogen sulphide is partially combusted with air to produce hydrogen sulphide, sulphur dioxide and sulphur. The remaining hydrogen sulphide and sulphur dioxide are converted to sulphur in one or more catalytic reactor stages. The sulphur is condensed out in a condenser separator by exchange with boiler feed water to produce L.P. steam. The condensed molten sulphur flows to the sulphur slating unit. The sulphur-containing tail gases flow to the SCOT tail-gas processing unit.

Area 1620 Sulphur Slating

Liquid sulphur from the Claus Unit is cooled, slated in a conventional sulphur slating unit and stockpiled. The sulphur product is bright yellow product, 99.8 + wt% sulphur, with less than 1 ppm H_2S content.

Area 1630 SCOT Tail Gas Processing

The SCOT process is a proprietary process licensed by the Shell Development Company. The feed gas is mixed with a reducing gas, preheated, and sent to a reactor where the sulphur compounds are catalytically hydrogenated to hydrogen sulphide. The hydrogen sulphide-rich gas is treated in a recovery section using a selective amine solvent. The recovered hydrogen sulphide is recycled to the Claus sulphur recovery unit The vent gases from the hydrogen sulphide absorber flow to an incinerator for incineration of remaining sulphur compounds in the gas to sulphur dioxide. The stack effluent has a sulphur dioxide content below the environmental regulatory limit.

.7 Section 1700 Gas Liquor Separation and Ammonia Recovery

The Koppers-Totzek gasification process takes place at a very high temperature (1600°C). Consequently, the produced raw gas does not contain condensible hydrocarbons, tar or phenols. Ammonia is present in the wastewater only in trace quantities. No gas liquor and ammonia recovery separation facilities are required for the Koppers-Totzek technology.

.8 Section 1800 Effluent Treatment

Area 1810 Fines Settling and Wastewater Treatment

Fines recovered from the washer cooler and electrostatic precipitators are settled out in a clarifier. The overflow water is recycled as wash water via the cooling tower. Settled clarifier solids are pumped to a settling pond where the solids settle as a sludge and the discharge water reused as ash handling water.

Area 1820 Solids Disposal

The molten ash from each gasifier is quenched and granulated in a water bath and removed by a continuous scraper system. This ash plus the sludge from the wastewater treatment system are transported by conveyor approximately 1 km to the mine site and disposed as landfill.

.9 Section 1900 Utilities, Services, Buildings and Offsites

Area 1910 Steam Generation and Distribution

Three levels of steam are used as follows:

- High Pressure (HP) superheated steam at 10800 kPa, 480°C.
- Medium Pressure (MP) saturated steam at 3500 kPa.
- Low Pressure (LP) saturated steam at 550 kPa.

There is a net steam requirement of 10900 t/d of 10800 kPa superheated steam which is generated in a steam boiler. The steam boiler is fired using as-received lignite from raw material storage supplemented by purge gas from the methanol unit. The boiler is supplied as a package unit consisting of pulverizer, lignite feeder, forced and induced draft fans, superheater, soot blowers, electrostatic precipitators, ducts, stack and all required instrumentation. Flue gas desulphurization facilities are not required as the sulphur compounds emission rates are well below the regulatory limits.

High pressure superheated steam is also generated in the gasifier waste heat boiler and the combined amount is used in backpressure and condensing turbine drivers for the compressors in the raw gas compression unit, air separation plant and methanol synthesis unit. The MP steam requirements are provided by heat recovery in the methanol converter and exhaust from back pressure turbines. The steam users consist of the CO shift, acid gas removal and sulphur recovery units. The L.P. steam requirements are provided by gasifier jacket steam, heat recovery in the CO shift unit and exhuast steam from backpressure turbines. The users consist of lignite drying, methanol purification and miscellaneous low pressure consumers such as heat tracing, utility stations and building heating.

Area 1920 Power Generation and Distribution

The power consumption for the Koppers-Totzek Base Case and Options is as shown in Table 3-4. For the consumption of 9.22 MW it is not considered economical to generate power on site. Thus power is purchased from Ontario Hydro and supplied from the existing 115 kV line in Onakawana, Ontario.

Power distribution within the plant is provided by standard substations, switchgear and motor control centres etc.

Area 1930 Cooling Water System

The cooling water recirculation rates for the overall plant for the base case and options are listed on Table 3-4.

A design Δt of 20°C was used for estimating cooling water loads. An induced draft cooling tower is required to perform the cooling duty. Blowdown, evaporation and drift losses are estimated to be 4 percent. Make-up water to the cooling tower is provided by the condensate blowdown from the condensate recovery system, clarified quench water and filtered raw water.

Area 1940 Raw Water

For the Phase 1 study, it is assumed that raw water is readily available from a river adjacent to the plant site. Also, raw water storage is considered unnecessary. Raw water used in the plant is passed through a flocculating clarifier and sand filter to remove suspended solids prior to distribution to the cooling water, service water and to the Boiler Feed Water System. The firewater pumps are also included in the raw water area.

Area 1950 Boiler Feed Water

The boiler feed water treatment plant treats the make-up raw water and recovered condensate. The system consists of a package BFW demineralizer, ion exchange unit, deaerators, chemical feed systems and associated pumps.

Area 1960 Fire Water

The fire water system consists of a firewater loop from the pumps in the raw water area around the process units and offsites. Fire hydrants and monitors are included.

Area 1970 Buildings

Buildings include the following:

- Administration Building
- Control Building and Laboratories
- Change House and Cafeteria
- Maintenance and Warehouse Buildings
- Gasifier Building
- Compressor Shelters.

Area 1980 Offsites

The offsites facilities consist of the following:

Raw Material Receiving and Storage

A transfer conveyor (not within plant battery limits) delivers lignite from the mine storage slab. The slab is designed such that the water and fines run-off is collected and discharged to effluent treatment. A drum reclaimer and transfer conveyor are provided to transfer the stored lignite to the lignite preparation area.

Product Storage

Thirty-days production of methanol storage is provided in cone-roof API tanks. Two 40000 m^3 (250,000 BBL) tanks are required.

Catalyst and Chemicals Storage

Storage facilities for the catalysts for methanol synthesis, water treatment chemicals and miscellaneous chemicals are provided.

Interbattery Pipes and Piperacks

This system includes all piping and pipe racks, bridges and foundations which support the piping between process units and from the units to the storage area.

Product Shipping

A loading pump and four-position rail loading rack and piping are provided for methanol shipping.

Flare

Two separate discharge flare and relief piping systems - a high pressure and a low pressure - are provided for the gasification section and the process units.

3.2.3 Option 1 Facilities Description

.1 Section 1101 Lignite Preparation

The general description of facilities is similar to the base case except that the capacity of the plant is approximately one quarter that of the base case.

.2 Section 1201 Gasification

Two gasifiers, one operating and one spare, are required for Option 1. A spare gasifier is considered necessary as methanol production would reduce to 52% of design capacity if only one gasifier is provided and this gasifier is shut down. Gasification facilities are as described for the base case, Section 1200.

.3 Section 1301 Syngas Compression and Acid Gas Removal

Area 1311 Syngas Compression

The gasifier generated raw gas is compressed to about 3000 kPa in steam turbine driven compressors.

Area 1331 Acid Gas Removal

The Rectisol unit is required for treatment of the gasifier generated syngas only. The capacity of the unit is approximately one quarter of the base case.

.4 Section 1401 Methanol Synthesis & Purification

Facilities are similar to the base case except that capacity is larger at 2648 t/d.

.5 Section 1501 Air Separation and Steam Methane Reforming

Area 1511 Air Separation Plant

Facilities are similar to the base case except that the capacity is 654 t/d 99.5% purity oxygen.

Area 1521 Steam Methane Reforming

Natural gas is delivered to the plant from the existing Trans-Canada Pipeline in a new pipeline, approximately 180 km long. The connection to the Trans-Canada Pipeline is made in Smooth Rock Falls, Ontario. The new pipeline follows the route of Highway 807 as far as Killoran, Ontario. From Killoran to Onakawana the Ontario Northland Railway route is followed. Natural gas pressure at the plant gate is 860 kPa. Natural gas is first desulphurized in zinc oxide guard beds. The desulphurized gas is compressed from 860 kPa to 3600 kPa in an electrically driven compressor, mixed with MP superheated steam, and reformed in catalyst filled tubes at approximately 760-870°C, using a nickel based catalyst.

The following reactions take place:

$$CH_4 + H_2O \qquad 3H_2 + CO$$
$$CO + H_2O \qquad CO_2 + H_2$$

The reformer feed rate is 980 t/d of natural gas. Reformed gas is combined with the syngas from the gasifier and fed to the methanol synthesis unit. Approximately 61 percent of the reformer fuel is provided by natural gas and the remaining 39 percent by purge gas from the methanol synthesis unit.

.6 Section 1601 Sulphur Recovery

Sulphur recovery, sulphur slating and tail gas processing facilities are as described in the base case. Sulphur recovery for Option 1 is 11 t/d.

.7 Section 1701 Gas Liquor Separation

These facilities are not required to the Koppers-Totzek technology.

.8 Section 1801 Effluent Treatment

Facilities are similar to the base case except that the capacities are lower due to the reduced number of gasifiers.

.9 Section 1901 Utilities, Services, Buildings and Offsites

Refer to Table 3-4 for the utility requirements for Option 1.

The net requirement to balance the steam system is 2329 t/d of 10800 kPa, 480°C superheated steam, generated in a lignite fired boiler. Superheated steam at the 10800 kPa level is also generated in the gasifier waste heat boiler and heat recovery facilities in the steam reforming unit. Refer to the base case Area 1910 for the description of the steam system.

Facilities for utilities, buildings and services are as described in the base case. Capacity of methanol storage facilities will be approximately 30 percent larger due to the higher methanol production.

3.2.4 Option 2 Facilities Description

.1 Section 1102 Lignite Preparation

Facilities are similar to the base case except the capacity is approximately one half of the base case.

.2 Section 1202 Gasification

Two operating gasifiers are required for Option 2. A spare gasifier is not required since methanol synthesis can continue at 91 percent of design capacity in the event one gasifier is shut down; steam reformed syngas bypasses the H.T. shift and PSA units and combines directly with the gasifier generated syngas to produce methanol. Gasification facilities are as described in the base case Section 1200.

.3 Section 1302 Syngas Compression and Acid Gas Removal

Facilities required are as described for Option 1 although at a higher capacity.

.4 Section 1402 Methanol Synthesis and Purification

Facilities are as described in the base case. Methanol production for Option 2 is 2533 t/d.

.5 Section 1502 Air Separation, Steam Methane Reforming, Partial Oxidation, Pressure Swing Adsorption

Area 1512 Air Separation Plant

Facilities are similar to the base case except that the capacity is 1308 t/d of 99.5 percent purity oxygen.

Area 1522 Steam Methane Reforming

Facilities are as described in Option 1 except that the reformer feedstock requirements are 790 t/d of natural gas. Reformer fuel is provided by fuel gas generated within the complex supplemented by 71 t/d of imported natural gas.

Area 1542 H.T. Shift & Pressure Swing Adsorption (PSA)

Reformed gas from Area 1522 passes through an H.T. Shift Conversion unit. Shift Conversion facilities are as described for Area 1320.

The PSA unit processes shifted synthesis gas to yield a high purity H_2 product gas suitable for methanol production. Overall hydrogen recovery is about 90 percent.

The unit uses a four-bed adsorption system with one bed on adsorption and three beds in various stages of regeneration. Shifted synthesis gas is fed to the on-stream adsorption bed where light hydrocarbons, water, carbon monoxide and carbon dioxide molecules are adsorbed. Hydrogen in the feed gas passes through the adsorption bed to form the recovered hydrogen stream. When the adsorption capacity of the bed is depleted, the feed gas is switched to a regenerated adsorber and the depleted adsorber is regenerated.

Regeneration of the adsorption bed is accomplished in three stages - depressurization, stripping with recovered hydrogen and repressurization to adsorption pressure. During the depressurization and stripping, hydrocarbons, water, carbon monoxide and carbon dioxide are desorbed, discharged at low pressure and routed to the fuel gas system.

The hydrogen product combines with the syngas from the acid gas removal unit and the combined gas is sent to the methanol synthesis unit for methanol production.

.6 Section 1602 Sulphur Recovery

Facilities are as described in the base case. Sulphur recovery for Option 2 is 23 t/d.

.7 Section 1702 Gas Liquor Separation

These facilities are not required for the Koppers-Totzek technology.

.8 Section 1802 Effluent Treatment

Facilities are similar to the base case but the capacities are lower due to the reduced number of gasifiers.

.9 Section 1902 Utilities, Services, Buildings & Offsites

The utility requirements for Option 2 are given on Table 3-4.

The net requirement to balance the steam system is 4136 t/d of HP superheated steam. The steam systems are generally as described for the Base Case and Option 1.

Facilities for utilities, buildings and services are as described in the base case.

3.2.5 Option 3 Facilities Description

The process scheme for Option 3 differs from Option 2 in that a water electrolysis plant replaces the natural gas reforming, H.T. Shift and PSA complex for hydrogen production. The process design for Option 3 has been optimized to include a CO shift unit and an electrolysis plant to supply the remaining hydrogen to correct the hydrogen deficiency in the syngas for methanol synthesis. To provide sufficient oxygen for gasification an air separation plant of 1204 t/d oxygen capacity is provided to supplement oxygen produced from the electrolysis plant.

.1 Section 1103 Lignite Preparation

Facilities are similar to the base case except the capacity is approximately one half of that of the base case.

.2 Section 1203 Gasification

Three operating gasifiers are required for Option 3. No spare gasifier is required since the methanol production would only be reduced to 75% of design capacity in the event one gasifier is shut down. Facilities are as described in the base case Section 1200.

.3 Section 1303 Syngas Compression, Shift Conversion & Acid Gas Removal

Area 1313 Syngas Compression

The syngas produced in the gasifiers is compressed to approximately 3000 kPa in steam turbine driven compressors.

Area 1323 Shift Conversion

The syngas is shifted in facilities as described in the base case, Area 1320.

Area 1333 Acid Gas Removal

The gasifier produced raw gas is treated in a Rectisol unit to remove H_2S and CO_2. The facilities are as described in the base case, Area 1330.

.4 Section 1403 Methanol Synthesis and Purification

The facilities are identical to the base case. Methanol production is 2010 t/d.

.5 Section 1503 Air Separation and Water Electrolysis

Area 1513 Air Separation

A 1204 t/d oxygen plant is required. Facilities are as described in the base case, Area 1510.

Area 1553 Water Electrolysis

Electrolysis of water is achieved by passing direct electric current between two electrodes immersed in an alkaline or acid electrolyte. Hydrogen and oxygen are generated at opposite electrodes according to the following reactions.

For alkaline electrolytic cell:

$$4e + 4H_2O \longrightarrow 2H_2 + 4OH \text{ cathode}$$
$$4\,OH - 4e \longrightarrow O_2 + 2H_2O \text{ anode}$$

There are three basic designs of electrolysers commercially available; these are the unipolar or tank electrolyser, the bi-polar or filter press type and the solid Polymer Electrolyte (SPE) electrolyser. For this study only the unipolar electrolyser was considered due to its simplicity, reliability, long life, low cost, and advanced status of commercialization and technological development. The water electrolysis plant consists of four essential components: the electric power conditioning equipment, the water demineralization unit, the electrolysis cell modules, and the gas conditioning and electrolyte recovery equipment for electrolyte systems. Water demineralization facilities are included in the utilities area with boiler feed water treatment.

The unipolar electrolysis cells are assembled in modules. Each cell consists of a series of electrodes, anodes and cathodes which are alternately suspended vertically and parallel to each other in a tank containing the electrolyte. The electrodes are connected in parallel allowing the tank to operate across a low voltage d.c. supply. Thus, each electrode has one polarity relative to the other electrode; hence the description - unipolar. An asbestos diaphragm surrounds alternate electrodes, normally the cathodes, and prevents the passage of gas from one electrode compartment to another. The diaphragm, although impermeable to gas, must be permeable to the electrolyte in order to offer the least resistance to electron flow through the electrolyte.

The unipolar cell has the following features:

Plant Size:	floor area 0.15 m^2/Nm^3 H_2/h
Cell Size:	volume $2m^3$, capacity 40 Nm^3 H_2/h
Energy Efficiency:	83%
Cell Voltage:	1.8 volts
Power Consumption:	4.4 kWh/Nm^3 H_2 or 117 kWh/1,000 scf H_2

Current Density:	250 mA/cm^2
Temperature:	70°C
Operating Mode:	fully-automated
Operating Flexibility:	rapid start-up to full load and shut-down to no load
Operating Reliability:	nominal maintenance, safe, long-life
Electrolyte:	25% KOH solution
Hydrogen Purity:	99.9%
Oxygen Purity:	99.7%

The electrolysis plant produces 95 t/d of hydrogen and 758 t/d of oxygen. Compression facilities are provided to deliver hydrogen to methanol synthesis at the make-up syngas compressor inlet pressure of 2000 kPa.

.6 Section 1603 Sulphur Recovery

Facilities are as described in the base case, Section 1600. Sulphur recovery for Option 3 is 35 t/d.

.7 Section 1703 Gas Liquor Separation

These facilities are not applicable to the Koppers-Totzek technology.

.8 Section 1803 Effluent Treatment

Facilities are identical to the base case although the capacities are lower.

.9 Section 1903 Utilities, Services, Buildings and Offsites

The utility requirements for Option 3 are given on Table 3-4.

The net steam requirement to balance the steam system is 8475 t/d of HP steam. The steam producers and users are generally as described for the base case with added consumers being the electrolysis plant hydrogen and oxygen compressor steam turbine drives.

Facilities for utilities, buildings and services are as described in the base case.

3.2.6 Overall Thermal Efficiency

The overall thermal efficiencies for the base case and options are shown in Table 3-8.

3.3 LURGI GASIFICATION TECHNOLOGY

3.3.1 Introduction

The Lurgi fixed bed process is one of the most established commercially proven gasification technologies. It has been continuously developed since the 1930's resulting in the gasification process being used in a total of 16 plants comprising approximately 100 gasifiers. The most significant commercial usage of Lurgi gasifiers has been in South Africa where SASOL I and II plants have a total of 52 gasifiers in operation and the SASOL III plant presently under construction will have 36 gasifiers in service.

There are no commercial plants processing lignite in Lurgi gasifiers. A major commercial venture in the U.S.A. currently in construction is the Great Plains Project in North Dakota, which will convert 22,000 t/d lignite to SNG utilizing Lurgi gasification technology.

The Lurgi process displays the following advantages and disadvantages when processing lignite:

Advantages:

(i) The syngas produced contains sufficient hydrogen that shift conversion is not required for methanol synthesis.

(ii) The process has been commercially operated for 50 years. The experience gained has resulted in improving operating and maintenance characteristics of the gasifier.

(iii) The Lurgi process can accept feedstock with a moisture content as high as 36 percent, though 30 percent is preferred. This reduces the duty of the drying equipment considerably.

(iv) The Lurgi process operates at a high pressure (3 MPa) resulting in low syngas compression requirements for methanol synthesis.

Disadvantages:

(i) The Lurgi gasifier operates in the range of 750-800°C which is generally lower than the fluidized bed and entrained bed gasifiers. This results in the production of a gas liquor containing tars, tar oils, phenols, and ammonia. Gas liquor recovery contributes to a high cost for downstream processing.

(ii) The Lurgi gasifier does not operate under slagging conditions hence an excess of steam is used to prevent the ash from melting. Consequently, there are large quantities of wastewater to be recovered and subsequently treated. Lurgi have developed a slagging gasifier in collaboration with British Gas but it has not been proven with lignites.

(iii) The syngas produced by the Lurgi process has a high methane content (10-15% v/v) which acts as an inert in the methanol synthesis reaction. For optimum yield the methane must be reformed and recycled to the main syngas stream.

(iv) The feedstock size range for the Lurgi gasifier is restricted such that fines cannot be processed. This is a serious drawback when processing lignite, since lignite tends to decrepitate when submitted to size reduction and drying operations. For the Great Plains Project, it is estimated that 35% of the feedstock will be rejected as fines and shipped to a nearby power station for use as fuel.

The process facilities for the base case and the optional case for a lignite-to-methanol plant are described as follows:

- The block flow diagram describing the base case (steam reforming of methane) is shown on Drawing 200-43-01.

- The block flow diagram describing the optional case (partial oxidation of methane) is shown on Drawing 200-43-02.

- Material balance information showing main stream flow rates is shown on these drawings.

3.3.2 Facility Description - Base Case

.1 Section 2100 - Lignite Preparation

Run-of-mine lignite is crushed to 50 mm x 5 mm size in primary and secondary crushers. The lignite is then dried from 45 percent moisture to about 30 percent for gasification. It is assumed that 25 percent of the lignite will be rejected as fines below 5 mm size. Only 10 percent of fines are routed to the utilities section for steam generation while the remainder is sold at zero value to a utility. Lignite feed for gasification is transported by conveyor to storage silos.

.2 Section 2200 - Gasification

Area 2210 Gasifier and Gas Quench

A total of six gasifiers, five operating and one spare, are required. The gasifier is a pressure vessel with water jacket, in which the lignite moves counter-current to the gasification agent. Lignite is charged through the top by means of lock hoppers and a coal distributor. Oxygen (from the air separation plant) and steam, the gasification agents, are injected through a rotary grate into the gasifiers. The steam also serves to cool the ash thus maintaining it below its melting point. The ash is removed through the rotary grate and discharged via lock hoppers. The gasifiers operate at 3 MPa (430 psig).

Raw gas exiting the gasifiers is quenched and cooled to condense the tar, oil, phenols, ammonia and water from the gas stream. These components comprise the gas liquor which are treated in Section 2700, Gas Liquor Separation. The raw gas, saturated with steam is further cooled by generating steam in waste heat boilers, and passes to the Raw Gas Cooling Unit, Area 2220.

Area 2220 Raw Gas Cooling

The gas cooling unit is designed to further cool the raw gas in an efficient manner and remove heavier hydrocarbons and unreacted steam. The gas stream leaving the unit is treated in the Acid Gas Removal Section, Area 2330, while the condensed liquids are treated in the Gas Liquor Separation, Section 2700.

Process conditions at the Gasification (Section 2200) battery limits are:

Gas flow:	5.74×10^6 Nm^3/d (5613 t/d)	
Gas Pressure:	2900 kPa	
Gas Composition:	H_2	39.02
(mole %)	CO	15.73
	CO_2	32.72
	CH_4	10.88
	C_2H_6+	0.75
	Naphtha	0.19
	H_2S & COS	0.58
	N_2	0.13
		100.00
Molecular Weight:	21.9	
H_2/CO:	2.48	

.3 Section 2300 Syngas Compression and Treating

Area 2310 Syngas Compression

Syngas pressure at the gasification section battery limits is about 2900 kPa which is sufficient for the acid gas removal, hence no syngas compression is required at this stage of the process.

Area 2320 Shift Conversion

The molar ratio of H_2 to CO in the Lurgi gasifier raw gas is 2.48 which is adequate for methanol synthesis. Consequently shift conversion is not required.

Area 2330 Acid Gas Removal

The acid gas removal unit employs the Lurgi Rectisol process which is designed to remove H_2, COS, CO_2 and other impurities from the synthesis gas by means of methanol as a physical solvent. Because of the different solubilities of H_2S and CO_2 in methanol, these components can be recovered selectively. This permits concentration of the H_2S to sufficient levels for recovery of the sulphur in conventional Claus sulphur plants. After absorption of the gaseous impurities the methanol is regenerated and recycled for further absorption.

As the absorption capacity of methanol increases with decreasing temperature, a methanol temperature in the order of -40°C is maintained.

The Rectisol unit consists of three basic process units. The first unit is the prewash step where naphtha, unsaturated hydrocarbons and other high boiling point impurities are removed. Sulphur compounds such as H_2S and COS are selectively removed from CO_2. The H_2S and COS components are removed to less than 1 ppm while the CO_2 content is removed to approximately 5.6 percent by weight for methanol synthesis. The acid gas stream is routed to the Sulphur Recovery and Tail Gas Processing Units, Section 2600.

Exiting synthesis gas conditions are:

Gas flow:		3.876×10^6 Nm³/d (1958 t/d)
		mol %
Gas composition:	H_2	57.79
	CO	23.30
	CO_2	1.46
	CH_4	16.11
	C_2H_6	1.16
	N_2	0.18
		100.00
Molecular Weight:		11.3

.4 Section 2400 Methanol Synthesis and Purification

Area 2410 Methanol Synthesis

The methanol synthesis area consists of the synthesis gas compressors and the methanol synthesis unit. The design of the methanol synthesis unit is based on the Lurgi low pressure process. Synthesis gas, which consists of combined streams from the acid gas removal units and the steam reforming unit, is compressed to 5500 kPa, preheated and fed to the methanol converter. In the reactor synthesis gas is converted to methanol by the following reactions:

$$2H_2 + CO \qquad CH_3OH$$

$$3H_2 + CO_2 \qquad CH_3OH + H_2O$$

The Lurgi methanol converter is of the tubular type with the synthesis catalyst in the tubes and a coolant in the shell to maintain an isothermal reaction for methanol synthesis. The heat of reaction is removed by generating medium pressure steam in the converter shell. Synthesized methanol is condensed in downstream heat recovery systems and transferred to intermediate storage for feed to the Methanol Purification Area 2420. Purge gas consisting mainly of methane is split with a portion transferred to the steam reforming unit for conversion to hydrogen rich syngas. This syngas is recycled to the methanol synthesis unit. The remaining purge gas is used as fuel gas in the complex.

Area 2420 Methanol Purification

The purification unit is as described for the Koppers-Totzek process, Area 1420.

.5 Section 2500 Air Separation Plant and Steam Methane Reforming

Area 2510 Air Separation Plant

For the base case design, an air separation plant of capacity 994 t/d of 99.5 percent purity oxygen is employed. The plant is of the standard low pressure design as described for the Koppers-Totzek process (Area 1510). Sufficient gaseous nitrogen is produced for the following uses:

- Inert blanketing and purging requirements
- Stripping CO_2 from methanol liquid in Rectisol acid gas removal unit
- Instrument and plant "air".

Area 2520 Steam Methane Reforming

Purge gas from the methanol synthesis unit, containing approximately 80 percent by weight of CH_4, C_2H_4 and C_2H_6, is routed to a steam reforming unit. This gas, already desulphurized in the acid gas removal unit, is mixed with superheated steam and reformed in catalyst (nickel-based) filled tubes at approximately 760-870°C and 3600 kPa. The following reactions take place:

$$CH_4 + H_2O \qquad\qquad 3H_2 + CO$$
$$CO + H_2O \qquad\qquad CO_2 + H_2$$

Approximately two-thirds of the purge gas to the reformer unit is used as feed and the remaining one-third used as fuel.

The reformed gas is recycled as feed to the methanol synthesis unit.

.6 Section 2600 Sulphur Recovery, Slating and Tail Gas Processing

Area 2610 Claus Sulphur Unit

The H_2S rich gas from the Acid Gas Removal unit (Area 2330) is sent to the Claus Sulphur unit. The facilities are as described in Section 1600 for Koppers-Totzek process. Approximately 49 t/day of elemental sulphur is recovered in molten form which is transferred to the sulphur slating unit (Area 2620).

Area 2620 Sulphur Slating

Liquid sulphur is cooled and slated in a conventional Sulphur Slating Unit and stockpiled. The sulphur product, 49 t/d, is bright yellow product, 99.8+ wt% sulphur, with less than 1 ppm H_2S content.

Area 2630 SCOT Tail Gas Processing

Off-gases from the Claus unit flow to the SCOT tail gas processing unit for further processing. This unit is described in Area 1620 for the Koppers-Totzek Process. The off-gases are treated, incinerated, and vented to atmosphere with a maximum SO_2 content of 830 g/m^3, well below the allowable emission limit.

.7 Section 2700 Gas Liquor Separation

Gas liquor contains tar, tar oil, naphtha and dissolved compounds such as phenols, ammonia, CO_2 and H_2S.

Gas liquor streams from the gas quench and gas cooling units are cooled and reduced in pressure. The CO_2 and flammable gases are flashed from the liquor and scrubbed with water for recovery of ammonia and incinerated in the steam boiler.

The liquor flows to a large separator from which tar, the component heavier than water, is removed. The remaining liquor overflows to a final separation stage consisting of two separators where the tar oil, which is lighter than water, is removed. The tars and tar oils are sent to intermediate storage and used as fuel for the steam boiler. The aqueous stream from the final separation stage flows to the Phenosolvan Unit for further treatment.

.8 Section 2800 Effluent Treatment

Area 2810 Fines Settling

Fines are not tolerated in the Lurgi gasification process and, hence, facilities are not required.

Area 2820 Solids Disposal

Ash is discharged from the gasifiers via an ash lock to screw classifiers. The classifier discharge is transferred by conveyor to the mine site for disposal.

Area 2830 Wastewater Treatment, Phenosolvan Unit and Ammonia Recovery

The plant is designed such that water is reused to the maximum extent. The most significant wastewater source is the aqueous stream from the gas liquor separation unit (Section 2700). Other sources of wastewater which must be treated before reuse as ash handling water make-up are as follows:

- Cooling tower blowdown
- Paved area drainage
- Ash dewatering
- Methanol Synthesis and Purification process water

These streams are stored in a pond and treated by oil separation, flocculation and clarification. Recovered slop oil is used in the steam boiler as fuel. Sludge is sent to the ash disposal system.

The Phenosolvan unit is a process licensed by Lurgi Corporation. Incoming aqueous stream from the gas liquor separators is passed through gravel filters for removal of suspended matter, then through extractors where an organic solvent is used to extract phenols. The organic solvent is distilled and separated from the phenols and recycled to the extractors.

The crude phenol is sent to storage for subsequent incineration in the steam plant. Dephenolized process water is stripped in a deacidifier to remove dissolved CO_2 and H_2S. The H_2S is transferred to the sulphur recovery plant (Area 2610). The process water is distilled to remove ammonia and recycled as cooling tower make-up. Recovered 25 percent aqueous ammonia is further distilled to produce commercially pure anhydrous ammonia. Approximately 35 t/d anhydrous ammonia is recovered in the Phenosolvan unit, and transferred to storage in Offsites (Area 2950).

.9 Section 2900 Utilities, Services, Buildings and Offsites

Area 2910, 2911 Steam Generation and Distribution
(Base Case and Option 1)

Three levels of steam are used in the plant, 10800 kPa, 3500 kPa, and 550 kPa. A steam boiler is needed to generate approximately 8030 t/d of steam at the 10800 kPa and 3500 kPa levels. This steam is used for steam turbine driven compressors in the air separation and methanol synthesis units and for gasifier process steam. The remaining steam requirements are generated by heat recovery within the process units. The steam boiler is fired using lignite fines and recovered gas liquors such as tars, tar oils and phenols. The steam boiler will be supplied as a package consisting of pulverizer, lignite feeder, superheater, forced and induced draft fans, soot blowers, electrostatic precipitators, ducts, stack, and all required instrumentation. Flue gas desulphurization facilities are not required.

Area 2920 Power Generation and Distribution

Power consumption for the Lurgi cases is as shown in Table 3-5. For a maximum consumption of 20.6 MW it is not considered economical to generate power on site. Therefore, all the power requirement is purchased from Ontario Hydro. Power is supplied from the existing 115 kV adjacent to the plant in Onakawana.

Power distribution within the plant is provided by standard substations, switchgear, etc.

Area 2930 Cooling Water System

The cooling tower facilities are as described for the Koppers-Totzek process, Area 2930.

Make-up water to the cooling tower is composed of condensate blowdown from condensate recovery system, treated water from the Phenosolvan unit and filtered water from the raw water treatment system. Cooling water quality is maintained by pH control, and inhibitor and biocide additions to prevent corrosion and algae build-up.

Area 2940 Raw Water

For purposes of this study, it is assumed that raw water is readily available from a river or lake adjacent to the plant site. Also, raw water storage is considered unnecessary. Raw water used in the plant is passed through a flocculating clarifier and sand filter to remove suspended solids, prior to distribution to the cooling water, service water and to the boiler feed water system. The firewater pumps are included in the raw water area.

Area 2950 Boiler Feed Water

A boiler feed water treatment plant treats the filtered make-up raw water and recovered condensate. The system consists of a package BFW demineralizer, ion exchange unit, deaerators, chemical feed system and associated pumps.

Area 2960 Fire Water

The fire water system consists of a firewater loop from the pumps in the raw water area around the process units and offsites, and includes hydrants, monitors, etc.

Area 2970 Buildings

Buildings include the following:

- Administration Building
- Control Building and Laboratories
- Change House and Cafeteria
- Maintenance and Warehouse Buildings
- Gasifier Building
- Compressor Shelters

Area 2980 Offsites

The offsites facilities consist of the following:

Raw Material Receiving and Storage

A transfer conveyor (not included within plant battery limits) delivers lignite from the mine to a storage slab. The storage slab is designed such that the water and fines run-off is collected and discharged to effluent treatment. A drum reclaimer and transfer conveyor are provided to transfer lignite to the lignite preparation area.

Product Storage

Thirty-day production of methanol storage is provided in cone-roof API tanks. Two 40 000 m³ (250,000 BBL) tanks are required. Thirty-day ammonia production storage is provided in one sphere of capacity 1 000 m³.

Catalyst and Chemicals Storage

Storage facilities for the catalysts for methanol synthesis and steam methane reforming, water treatment chemicals and miscellaneous chemicals are provided.

Interbattery piping and piperacks

This system includes all piping and piperacks, bridges and foundations which support the piping between process units and from the process units to the storage area.

Product Shipping

A loading pump and four-position rail loading rack and piping are provided for methanol shipping. A loading pump and single-position rack are provided for ammonia shipping.

Flare

Two separate discharge flare and relief piping systems, one high-pressure and one low-pressure are provided for the gasification section and the process units.

3.3.3 Option A Facilities Description

This option entails the replacement of the steam reforming unit (Area 2520) by a partial oxidation unit (Area 2530). The facilities for Option A are as described for the base case except for the partial oxidation unit. A larger capacity air separation plant is required due to the oxygen feed to the partial oxidation unit and there are minor changes to utility requirements.

.1 Area 2540 Partial Oxidation

Purge gas from the methanol synthesis unit contains large quantities of methane and mixed C_2's. To increase methanol yield and hence reduce the fuel gas produced, part of this purge gas stream is recycled to the methanol synthesis unit through a partial oxidation process.

In this process purge gas and oxygen are preheated separately, combined and fed to partial oxidation reactors. The reactors are fire brick/refractory lined vertical pressure vessels. In the reactors, methane and mixed C_2's are partially combusted to hydrogen and carbon monoxide. Hot reactor effluent gas is then rapidly cooled in waste heat boilers producing high pressure saturated steam. Cooled gas from the waste heat boilers passes to boiler feedwater preheaters to recover additional sensible heat from the gas and finally to a water quench system. Water condensation and gas cooling occurs in the quench system. In addition, soot carried over from the reactors is scrubbed from the gas. Overhead gases from the quench system are recombined with the syngas from the acid gas removal system as feed to the methanol synthesis unit.

3.3.4 Thermal Efficiency

Overall thermal efficiencies for the base case and Option A are shown on Table 3-8.

3.4 WESTINGHOUSE GASIFICATION TECHNOLOGY

3.4.1 Introduction

The Westinghouse pressurized fluidized bed process has been operated on a pilot plant scale since 1975 and has logged over 10,000 hours of operation. A demonstration plant of 1200 t/d capacity planned for SASOL's facilities in Secunda, South Africa was recently cancelled due to prevailing economic conditions. As a result there is no schedule for commercialization at this time. There are four currently announced commercial projects utilizing the Westinghouse gasification technology in the planning stages as follows:

- Keystone Project in Johnstown, Pennsylvania consisting of a coal to methanol plant which is in the preliminary design stage.

- Gulf States Utility Project in Louisiana consisting of production of a medium BTU gas to fuel a Westinghouse combustion turbine which is in the preliminary design stage.

- Fiat/Ansaldo Project in Italy consisting of a coal gasification combined cycle system now in the project planning stage.

- New England Energy Park Project which would produce methanol, SNG and electric power in the feasibility stage.

Because of falling oil prices and escalating commercialization costs, Westinghouse is seeking buyers for its Synthetic Fuels Division.

The Westinghouse gasification technology has successfully gasified lignites on a pilot plant scale, particularly Texas lignite. A total of 257 hours of testing has been performed using Texas lignite during 1980 and 1981. North Dakota lignite has also been tested by Westinghouse but data have not been published due to proprietary considerations.

As the Westinghouse process has not been proven commercially an assessment of the various advantages and drawbacks of the technology is made with reference to literature sources.

Advantages:

(i) The gasifier is designed to operate at high pressures which allows a high capacity of about 1200 t/d per unit, and reduces syngas compression requirements for methanol synthesis.

(ii) Thermal efficiency is high in the heat recovery unit.

(iii) The gasifier accepts a wide range of raw material sizes, from 5 mm to fines.

Disadvantages:

(i) The syngas produced contains methane, which like the Lurgi process, must be reformed to optimize yields.

(ii) Materials handling problems have been reported with feed lignite and fines recycle systems.

(iii) There is no operating experience in the pilot plant at the pressures proposed for a commercial size gasifier.

Process facilities for the base case and four options are described on the following block diagrams.

- Drawing 300-43-01 describes the base case for a lignite to methanol plant utilizing lignite as a raw material only with shift conversion and steam reforming of the methane rich purge gas.
- Drawing 300-43-02 describes Option 1 for the co-processing option of lignite and natural gas.
- Drawing 300-43-03 describes Option 2 for the steam methane reforming/H.T. Shift/PSA co-processing scenario.

- Drawing 300-43-04 describes Option 3 for the electrolysis case producing pure hydrogen to correct the syrgas H_2/CO ratio. Electrical power requirements for the plant are purchased from Ontario Hydro.

- Drawing 300-43-05 describes a process similar to Option 3 except that power is co-generated in a combined cycle gasification unit.

3.4.2 Facilities Description - Base Case

.1 Section 3100 Lignite Preparation

Run-of-mine lignite is received in lump form at 45 percent moisture content. Lignite is reduced to size minus 6 mm x 0 size with no more than 10 percent passing 150 micron (100 mesh) in impact mill crushers. It is then dried to a free-flowing condition, recommended by the process developer at approximately 15 percent moisture content in steam tube rotary driers. The lignite is fed by conveyor to storage bins for feeding to the gasifiers.

.2 Section 3200 Gasification System

The gasification system consists of the following areas:

Area 3210	Gasifier including lignite feed lockhopper system and cyclone separators
Area 3220	Raw Gas Cooling
Area 3230	Quench System
Area 3240	Ash Collection

Four gasifier trains each containing lignite feed, raw gas cooling, quench and ash collection facilities are required. The four trains provide an excess capacity of 30 percent, thus affording the opportunity to run at almost full design capacity with three gasifiers on line. This allows shutdown periodically of one gasifier for maintenance and/or inspection. Each gasifier can process 1250 tonne/day of beneficiated lignite.

Area 3210 Gasifier

The lignite feed system consists of a surge bin arranged in series with the two lockhoppers for each gasifier. Lignite is repeatedly batched at atmospheric pressure from the surge bin into the first pressurization lockhopper. This lockhopper is pressurized in a batchwise manner to feed lignite into the second lockhopper. The second lockhopper operates at a constant pressure which allows pneumatic conveying of the feed to the gasifier. Recycle raw gas is used as a transport medium to convey the lignite feed continuously from the feed lockhopper to the gasifier.

Each fluidized bed gasifier is a 4.1 m O.D. refractory lined carbon steel pressure vessel. Operating pressure is 3100 kPa. The lignite, steam and oxygen enter through nozzles located in the ash separator at the bottom of the vessel. The initial combustion results in formation of a char which forms the gasifier bed material. The char bed is fluidized by incoming feeds and recycle fines and gas. Carbon in the char reacts with the steam to yield carbon monoxide and hydrogen. The volatile materials are cracked and gasified with steam to produce hydrogen, carbon monoxide and a little methane. Ash particles agglomerate, defluidize and gradually drop to the char ash separator region of the gasifiers and are removed in dry form by a rotary feeder. The raw gas produced passes upward through the increased diameter freeboard section which reduces particle entrainment and allows additional time for the reactions to proceed to equilibrium.

Exiting the gasifier the raw gas passes through the particulate recovery system consisting of two cyclones in series. The entrained particles collected in these cyclones are recycled to the gasifier using recycle gas as the transport medium.

A summary of the important mechanical aspects and process conditions in:

Gasifier operating conditions	
Pressure	3100 kPa
Temperature	950°C
Size of gasifiers	4.1 m O.D.
Number of gasifiers	4
Total coal feed to gasifiers (15% moisture)	3900 t/d
Gasification agent consumption	
Steam (3450 kPa, 400°C)	668 t/d
Oxygen (98.5 mole %)	1671 t/d
Total Raw Gas Flow	5492 t/d

Area 3220 Raw Gas Cooling

This system performs a heat recovery function as well as cooling the gas for further processing. The system consists of a feedwater heater, raw gas heat exchanger evaporator and steam superheater. Superheated steam is delivered at approximately 4200 kPa and 425°C.

Area 3230 Raw Gas Quench

This system is designed to cool the raw gas leaving the raw gas cooling system and to remove any remaining particulate matter entrained in the gas. The raw gas is intimately contacted with water in a venturi scrubber to adiabatically humidify and cool it.

The wet stream enters the particulate scrubbing quench unit where it is further cooled by direct countercurrent contact with water. A slurry is continuously drained to the pressure filtration system. The filtrate is recycled via the quench water filtration system.

The clean raw gas leaves the gasifier battery limit to downstream processing units. A small slipstream is withdrawn for use as a transport medium in the gasification system.

A summary of the process parameters at the gasification section battery limits is as follows:

Raw Gas Composition (mf):

	mol %
H_2	27.9
CO	53.4
CO_2	12.8
CH_4	4.5
H_2S & COS	0.7
N_2	0.7
Gas Flow Rate:	5492 t/d
Gas Pressure:	2860 kPa
Temperature:	165°C

.3 Section 3300 Gas Compression and Treating

Area 3310 Syngas Compression

As the pressure exiting the gasification section is sufficient for acid gas removal, compression facilities are not required at this stage of the process.

Area 3320 CO Shift Conversion

Refer to Koppers-Totzek technology Area 1320 for the description of the process. The composition of the syngas exiting the shift conversion unit is as follows:

	mol % (m.f)
H_2	43.7
CO	20.0
CO_2	31.8
CH_4	3.5
H_2S	0.5
N_2	0.5

Molar Ratio H_2/CO: 2.18. This ratio will be further increased when this syngas stream is combined with the reformed purge stream.

Area 3330 Acid Gas Removal

The Rectisol acid gas removal process is used, which is described for the Koppers-Totzek technology, Area 1330.

.4 Section 3400 Methanol Synthesis and Purification

Area 3410 Methanol Synthesis

Gas entering the methanol synthesis unit is a combination of syngas from gasification and syngas from steam methane reforming Area 3520. The composition of the gas is as follows:

	mol % (m.f)
H_2	64.7
CO	27.4
CO_2	2.8
CH_4	4.4
N_2	0.7

Area 3420 Methanol Purification

The purification unit is as described for the Koppers-Totzek technology Area 1420.

.5 Section 3500 Air Separation Plant and Steam Methane Reforming

Area 3510 Air Separation

The air separation plant design throughput is 1671 t/day of 99.5 percent purity oxygen. The description of the facilities is as given for the Koppers-Totzek technology Area 1510. Nitrogen is also used for the following:

- Transportation of lignite feed from lignite preparation to raw material storage silos
- Stripping CO_2 from methanol liquid in Rectisol Acid Gas Removal unit
- Miscellaneous blanketing and purging
- Instrument and plant air.

Area 3520 Steam Methane Reforming

The base case design involves a small steam reforming unit to convert methane (about 3.7% in the gasification unit syngas) in the purge gas to a syngas. The reformer total capacity is 154 t/d comprising 104 t/d of feed and 52 t/d of fuel. Steam for the reformer is generated in the convection section of the reformer furnace. A general description of the reformer facilities is as given for the Lurgi technology, Area 2520. The reformed gas combines with the gasifier syngas downstream of the acid gas removal unit.

.6 Section 3600 Sulphur Recovery, Sulphur Slating and Tail Gas Processing

The general description of the Claus Sulphur recovery unit, the sulphur slating unit and the SCOT tail gas unit is given for the Koppers-Totzek technology, Section 1600. About 48 t/d of sulphur product is recovered.

.7 Section 3700 Gas Liquor Separation

There is no gas liquor produced from the gasifier; hence separation facilities are not required.

.8 Section 3800 Effluent Treatment and Disposal

Area 3810 Fines Settling

The fines removed with the raw quench water are concentrated to a slurry in a clarifier. The overflow water is transferred to the ammonia recovery unit, Area 3720 for removal of ammonia and H_2S rich gas. The settled clarifier solids are pumped to the solids disposal area to be combined with the ash from the gasifier.

Area 3820 Solids Disposal

The agglomerated ash from the gasifiers is depressured through a lockhopper system using starwheel feeders and fed to a mixing unit where it is combined with the fines from fines settling. The combined ash is discharged by a conveyor to the mine as landfill.

Area 3830 Ammonia Recovery

Sour water streams from the gasification quench system, fines settling and the shift conversion unit are transferred to the ammonia recovery unit. A sour water WWT process developed by Chevron Research Co. is utilized. The unit removes hydrogen sulphide as a gas to a level of 5 mg/l H_2S in the water stream. Ammonia is removed as anhydrous ammonia and the resulting water stream contains 50 mg/l ammonia. Sour water is collected in an interbattery storage tank and pumped to the hydrogen sulphide stripper. Hydrogen sulphide is stripped overhead, water washed and sent to the sulphur plant, Area 1620. The water stream from the stripper containing ammonia and some hydrogen sulphide is fed to the ammonia stripper. The resulting ammonia and H_2S stream is scrubbed to remove the H_2S and the vapour is compressed and liquefied to produce 6.9 t/d of anhydrous ammonia. The ammonia is transferred to product tankage for sale as a by-product.

Area 3840 Wastewater Treatment

Aqueous streams containing traces of hydrocarbon from the following areas are collected and treated:

- paved area drainage,
- methanol synthesis and purification.

These streams are treated in oil separation, flocculation and clarification units. Sludge is sent to the solids disposal system.

.9 Section 3900 Utilities, Services, Buildings & Offsites

Area 3910 Steam Generation

The base case energy balance indicates that 4000 t/d of steam is required in excess of onsite waste heat recovery systems. This steam is generated at 10800 kPa and 480°C in a lignite fired steam boiler. Steam is produced at four pressure levels and is used as follows:

10800 kPa:	Air separation plant compressors.
4200 kPa:	Syngas make-up and recycle compressors in methanol synthesis unit.
3500 kPa:	Gasifiers, CO shift and reformer process steam requirements and in acid gas removal unit.
550 kPa:	Lignite drying and methanol purification unit; low pressure users such as steam tracing, building heating, etc.

Area 3920 Power Generation and Distribution

Power consumption for the base case and options is shown on Table 3-6. The electrical power requirements are purchased from Ontario Hydro. Power distribution within the plant is provided by standard sub-stations, switchgear, etc.

Areas 3930, 3940, 3950, 3960 Cooling Water, Raw Water Boiler Feed Water & Firewater

Refer to Table 3-6 for the consumption of these utilities. The general description of the facilities is as given for the Koppers-Totzek technology.

Area 3970 Buildings

The description of the facilities provided is given for the Koppers-Totzek process, Area 1970.

Area 3980 Offsites

Facilities are provided for the following:

- Raw material storage
- Product Storage, (Methanol, Ammonia, Sulphur)
- Interbattery piping and piperacks
- Flare system

The general description is as given for the Koppers-Totzek and Lurgi process.

3.4.3 Option 1 Facilities Description

Option 1 entails co-processing of lignite with natural gas to produce a syngas for methanol synthesis. A detailed description of facilities will not be given here as it is as provided for the Koppers-Totzek Option 1 and the Westinghouse base case. The areas involving significant changes from the base case are highlighted below.

.1 Section 3101 Gasification

One gasification train processing 993 t/day of feed lignite (15% moisture) is provided. A spare gasifier is required since the plant can operate at only 50 percent design methanol capacity on the syngas produced from steam methane reforming should the operating gasifier be shutdown.

.2 Section 3401 Methanol Synthesis and Purification

Methanol capacity is 2216 t/d for Option 1 (2227 t/d fuel grade methanol).

.3 Section 3521 Steam Methane Reforming

A reforming unit processing 737 t/d of natural gas is provided. Approximately 65 percent of the reformer fuel is provided by natural gas and the remaining 35 percent by purge gas from methanol synthesis.

.4 Area 3911 Steam Generation

High pressure steam generated in sufficient quantity by the gasifier waste heat boiler and natural gas reforming waste heat recovery systems to meet the requriements of the steam turbine driven compressors in the air separation plant and methanol synthesis unit. A lignite fired steam boiler of capacity 1687 t/d is required to balance the medium (3500 kPa) and low pressure (550 kPa) steam systems.

3.4.4 Option 2 Facilities Description

Option 2 entails production of pure hydrogen to correct the hydrogen deficit in the syngas from the gasifiers. The hydrogen is produced by steam reforming the natural gas, shifting the syngas, and passing the resultant syngas through a PSA unit. As the plant facilities have been described in discussion of other technologies, the base case and Option 1 of the Westinghouse technology, only areas of significant change are highlighted below.

.1 Section 3102 Gasification

Two operating gasifiers processing 1985 t/d of lignite feed (15% moisture) are provided. A spare gasifier is not required since in the event of one gasifier shutdown the PSA unit can be bypassed with methanol synthesis being performed on syngas produced from reformed natural gas and syngas produced from the remaining gasifier. This results in a 12 percent reduction of design capacity.

.2 Section 3402 Methanol Synthesis and Purification

Fuel grade in methanol production is 2144 t/d.

.3 Section 3522 Steam Methane Reforming

A reformer unit processing 641 t/d natural gas is provided. Reformer fuel is provided utilizing purge gas from the methanol synthesis unit.

.4 Section 3912 Steam Generation

As described for Option 1, the high pressure steam requirements are provided by heat recovery within the complex. There is a requirement to provide a lignite/fuel gas fired steam boiler to generate 2636 t/d of medium and low pressure steam to balance the system.

3.4.5 Option 3 Facilities Description

Option 3 entails using a water electrolysis plant to produce hydrogen for syngas H_2/CO ratio adjustment. Like the Koppers-Totzek Option 3, the design has been optimized to include a partial shift conversion of the gasifier syngas and a downscaled electrolysis plant to provide the bulk of the hydrogen. As the plant facilities have been described in previous sections, only significant changes or clarifications are highlighted below.

.1 Section 3103 Gasification

Two operating gasifiers processing 2878 t/d of feed lignite are provided.

.2 Section 3323 Shift Conversion

A shift conversion unit is provided to shift approximately 75 percent of the syngas from the gasification section with 25 percent bypassing the unit.

.3 Section 3350 Electrolysis Plant

An electrolysis plant with a total power consumption of 210 MW is required. Steam turbine driven compressors are provided to deliver the produced hydrogen and oxygen to the methanol synthesis plant and gasifiers respectively.

.4 Section 3903 Utilities

Area 3913 Steam Generation

The large compressors in the air separation plant, methanol synthesis unit and electorlysis plant are driven by steam turbines. There is a requirement for a steam boiler to generate 2891 t/d of 10800 kPa, 480°C steam to balance the steam system. The medium and low pressure steam requirements are similar to those described for Option 1 and are provided by exhaust from the backpressure turbines and the methanol converter.

3.4.6 Option 4 Facilities Description

Option 4 entails using a lignite gasification combined cycle system to provide the electricity required for Option 3 - electrolysis case. A separate three-train gasification section is provided for power generation. The desulphurized low BTU gas produced from the gasifiers is used as fuel in a combustor which generates the hot gas required to drive the turbines.

Except for the power generation system, all other facilities are identical to those described for Option 3. The gasification - power generation system is described below.

.1 Section 3904 Utilities

Area 3924 Power Generation

Total power consumption is 242 MW. This requirements is met by generating power on-site, using a lignite gasification combined cycle system. This system consists of three trains of gasifiers (2 operating plus one spare), raw gas cooling and quenching units, a sulphur removal unit, and a power generation unit including combustion turbine and steam turbine.

About 2680 t/d of dry lignite (15 percent moisture) is processed in the gasifiers utilizing air as the source of oxygen. The raw gas produced is passed through particulate removal, raw gas cooling (heat recovery) and quench scrubber units prior to sulphur removal in a Stretford sulphur plant.

The particulate-free, desulphurized clean gas enters the combustion turbine combustor, where it is burned with air to provide a hot gas that powers the turbine. The hot gas is expanded through the turbine and produces the work required to drive both the combustion turbine air compressor and its associated electric generator. The compressor is used to supply compressed air to the gasifiers.

Exhaust gas from the gas turbine is utilized in a final heat recovery steam generator to generate superheated steam for use in the main steam turbine system, which also has an associated electric generator.

The overall efficiency of this gasification combined cycle electric power generation system is about 40%.

3.4.7 Thermal Efficiency

The overall thermal efficiencies for the base case and options are shown on Table 3-8.

3.5 HIGH TEMPERATURE WINKLER GASIFICATION TECHNOLOGY

3.5.1 Introduction

The High Temperature Winkler process, like the Westinghouse process, is a fluidized bed gasification technology. The process is derived from the first generation atmospheric pressure low temperature Winkler gasifier. The low temperature Winkler gasifier has been in commercial use since 1926, with significant experience being gained using Rhenish brown coal, which is a lignite.

The low temperature Winkler process has displayed the following capabilities:

- Simple feedstock preparation, as a wide grain size range can be used.
- Low consumption of gasification agents, i.e. oxygen and steam.
- Satisfactory partial load behaviour over a wide performance range.
- Simple start-up and shutdown procedures.

The low temperature Winkler process, however has suffered from a number of drawbacks, the most significant being the following:

- Low capacity due to atmospheric pressure operation.
- High cost for compression of syngas.
- Low temperature operation causing incomplete carbon conversion and syngas contained partially oxygenated materials such as phenols and cresols.
- Fines carryover in the raw syngas.

Rheinische Braunkohlenwerke AG (Rheinbraun), who have operated many of the conventional atmospheric Winkler plants have developed the High Temperature Winkler fluidized bed technology in association with Uhde GmbH. A 30 t/d pilot plant has been in operation since 1978 at the site of their Wachtberg plant near Cologne, W. Germany. The advantages of the atmospheric process were still maintained with the following added features:

- Gasification under pressure aimed at increasing specific output and decreasing the compression energy demand. Conventional Winkler gasifiers operate slightly above atmospheric pressure while the developed version operates at 1100 kPa (160 psi) max.
- Gasification at elevated temperatures for improving the syngas quality and increasing the conversion rate, thus the High-Temperature Winkler process. Conventional Winkler units operate at below the melting point of ash (800°C for lignite, 1100°C for bituminous coal), therefore are best suited to highly reactive coals, i.e. lignite or reactive hard coals. The High-Temperature unit when processing lignite, has been operated as high as 1100°C. The melting point of the ash has been controlled by the addition of limestone to the feedstock.
- Increasing the carbon conversion rate by partially recycling flue dust to the gasifier.

The promising results which were achieved from the some 8000 hours of operation of the pilot plant prompted Rheinbraun's decision to install the first commercial size synthesis gas plant employing the High-Temperature Winkler process. The plant will be located at the site of Rheinbraun's lignite briquetting plant at Berranrath with the primary product being synthesis gas. The gas will be piped to the Union Kraftstoff plant, Wesseling for the production of methanol. Engineering work started in 1981, site preparation is underway and long delivery items have been ordered. The plant will ultimately comprise four units with the first one to be completed in 1984, which has a capacity of 41 000 Nm³/h of synthesis gas from 25 t/h of dry lignite (12% moisture). Methanol (14 t/h) will be produced from the synthesis gas.

The base case and options for lignite-to-methanol plants using the High Temperature Winkler process are described on the following block diagrams:

- Drawing 400-43-01 describes the base case for a lignite to methanol plant utilizing lignite as a raw material and CO shift conversion to correct the H_2/CO ratio.
- Drawing 400-43-02 describes Option 1 for the co-processing option of lignite and natural gas.
- Drawing 400-43-03 describes Option 2 for the steam methane reforming/H.T. Shift/PSA co-processing scenario.
- Drawing 400-43-04 describes Option 3 for the electrolysis case producing pure hydrogen to correct the syngas H_2/CO ratio.

3.5.2 Facilities Description - Base Case

.1 Section 4100 Lignite Preparation

The facilities are very similar to those described for Westinghouse technology, Section 3100. However, more drying is entailed as the process developer recommends a gasifier feed with 12 percent moisture.

.2 Section 4200 Gasification

The gasification section consists of the following:

- Inlet pressure locks
- Gasifiers
- Primary and Secondary Cyclones
- Waste Heat Boilers
- Ash Locks

Four gasifier trains are required, consisting of three operating and one spare. The dried lignite is brought to reaction pressure (1000-1100 kPa) in a lock hopper system. In the gasifier, the lignite reacts with the preheated gasification agent, composed of either steam/oxygen or air, which is introduced at several levels. The product gas, laden with dust, leaves the gasifier at the top. The dust laden gas passes through a waste heat boiler which generates superheated steam for use in the gasifier. The fines are removed in three trains (no spare is required), consisting of gas quench chambers, venturi scrubbers and separators. The saturated syngas exits the gasification section at about 1000 kPa and 40°C. The raw syngas composition on a dry basis is as follows:

	mol %
H_2	35.1
CO	47.4
CO_2	15.2
CH_4	1.1
H_2S & COS	0.7
N_2	0.5
	100.0

Average Molecular Wt: 21.2

Raw Gas flow: 5.616×10^6 Nm³/d

.3 Section 4300 Syngas Compression and Treating

Area 4310 Syngas Compression

The raw syngas is compressed in centrifugal compressors from 1000 kPa to 3000 kPa. Water in the gas is removed in the compressor intercoolers and aftercoolers.

Area 4320 Shift Conversion

Refer to Koppers-Totzek Area 1320 for a general description of the process. The exit gas composition is as follows:

	mol %
H_2	47.2
CO	19.8
CO_2	31.1
CH_4	0.9
H_2S + COS	0.6
N_2	0.4
	100.0

Molar Ratio H_2/CO: 2.38

Area 4330 Acid Gas Removal

The Rectisol acid gas removal process is used, for which the description is given in the Koppers-Totzek technology base case, Area 1330.

.4 Section 4400 Methanol Synthesis and Distilation

Area 4410 Methanol Synthesis

The methanol synthesis facilities are as described for the Koppers-Totzek technology, Area 1340. An exception to the Koppers-Totzek technology is that some of the purge gas is recycled to the gasifiers as a seal gas for the ash handling system. Excess purge gas is utilized as fuel for lignite drying and in the steam boiler. The composition of the synthesis gas is as follows:

	mol %
H_2	66.9
CO	28.0
CO_2	3.3
CH_4	1.3
H_2S + COS	0.5
	100.0

Area 4420 Methanol Purification

Refer to Koppers-Totzek technology Area 1420 for a description of the process.

.5 Section 4500 Air Separation Plant

Area 4510 Air Separation Plant

The air separation plant design throughput is 2200 t/day of 99.5 percent purity oxygen. The general description of the facilities is as given for the Koppers-Totzek technology Area 1510.

.6 Section 4600 Sulphur Recovery, Slating and Tail Gas Processing

The general description of these facilities is as given for the Koppers-Totzek technology, Section 1600. Approximately 56 t/d of 99.8+ percent pure sulphur is produced.

.7 Section 4700 Gas Liquor Separation

The High Temperature Winkler operates at a sufficiently high temperature to prevent condensible hydrocarbon formation, hence gas liquor separation facilities are not required.

.8 Section 4800 Effluent Treatment and Disposal

Areas 4810, 4820, 4830, 4840 Fines Settling, Solids Disposal, Ammonia Recovery, Wastewater Treatment

Refer to the Westinghouse technology, Section 4800 for a general description of these facilities, which are very similar for the High Temperature Winkler technology. About 12 t/day of anhydrous ammonia is produced in the ammonia recovery unit.

.9 Section 4900 Utilities, Services, Buildings and Offsites

Area 4910 Steam Generation

There is a net requirement of 7000 t/d, 10800 kPa, 480°C steam to balance the steam system. This steam is used exclusively in backpressure and condensing steam turbine drivers for the compressors in the air separation plant, raw gas compression and methanol synthesis units. The medium pressure steam (3500 kPa) requirements in the gasifier, CO shift acid gas removal, sulphur and ammonia recovery units are supplied by the gasifier waste heat boiler and methanol coverter heat recovery. The low pressure (550 kPa) steam requirements for lignite drying, methanol purification and miscellaneous users are supplied by the backpressure turbines exhaust and heat recovery in the CO shift unit.

Area 4920 Power Generation and Distribution

Power consumptions for the base case and options are shown on Table 3.7. The electrical power requirements are purchased from Ontario Hydro. Power distribution within the plant is provided by standard substations, switchgear, etc.

Areas 4930, 4940, 4950, 4960 Cooling Water, Raw Water, Boiler Feed Water and Fire Water

Refer to Table 3.7 for the requirements of these utilities. The general description of the facilities is as given for the Koppers-Totzek technology.

Area 4970 Buildings

The description of the facilities is as given for the Koppers-Totzek process, Area 1970.

Area 4980 Offsites

Facilities are provided for the following:

- Raw material storage
- Product storage (methanol, ammonia, sulphur)
- Interbattery piping and piperacks
- Flare system.

The general description is as given for the Koppers-Totzek and Lurgi processes.

3.5.3 Option 1 Facilities Description

Option 1 entails co-processing of lignite with natural gas to produce a syngas for methanol synthesis. A detailed description of facilities is not given here as it is as provided for the Koppers-Totzek Option 1 and the High Temperature Winkler base case. The areas of significant change from the base case are highlighted below.

.1 Section 4101 Gasification

One operating gasification train processing 1198 t/d of feed lignite (12% moisture) is provided. A spare gasifier is required since the complex can operate at only 48 percent of design capacity using syngas produced from the steam methane reforming unit.

.2 Section 4401 Methanol Synthesis and Purification

Methanol capacity is 2780 t/d for Option 1.

Area 4521 Steam Methane Reforming

A reforming unit processing 877 t/d of natural gas is provided. Approximately 77 percent of the reformer fuel is provided by natural gas and 23 percent by purge gas from the methanol synthesis unit.

Area 4911 Steam Generation

There is a requirement to provide a lignite fired steam boiler to generate 2445 t/d of high pressure superheated steam to balance the steam system. This steam produced in the boiler and superheated steam generated by reformer waste heat recovery meets the steam turbine requirements. The MP and LP requirements and producers are similar to those described for the base case.

3.5.4 Option 2 Facilities Description

Option 2 entails the production and supply of pure hydrogen to correct the hydrogen deficiency in the gasifier syngas. The hydrogen is produced by steam reforming the natural gas, shifting the syngas and passing the shifted syngas through a PSA unit. As similar plant facilities have been described for this and other technologies, only areas of significant change are highlighted below.

.1 Section 4102 Gasification

Two operating gasifiers processing 2397 t/day of lignite feed are provided. A spare gasifier is not required since in the event of one gasifier shutdown the PSA unit can be bypassed and methanol synthesis performed on the syngas produced from the steam methane reforming unit and the syngas produced from the remaining gasifier. This results in a 17 percent reduction design capacity.

.2 Section 4402 Methanol Synthesis and Purification

Methanol production is 2882 t/d.

.3 Section 4522 Steam Methane Reforming

A reformer unit processing 761 t/d natural gas is provided. About 77 percent of the reformer fuel requirement is provided by fuel gas from the methanol synthesis and PSA units with the remaining 23 percent by natural gas.

4. Section 4912 Steam Generation

There is a requirement to provide a steam boiler to generate 3786 t/d of high pressure steam to balance the steam system. The system is generally as described for Option 1.

3.5.5 Option 3 Facilities Description

Option 3 entails using a water electrolysis plant to produce hydrogen for syngas H_2/CO ratio adjustment. As discussed for Koppers-Totzek and Westinghouse technologies the design has been optimized to include a partial shift conversion of the syngas and a downscaled electrolysis plant to provide the bulk of the hydrogen. As the plant facilities have been generally described in previous sections, only significant changes or clarifications are highlighted below.

.1 Section 4103 Gasification

Two operating gasifiers processing 2397 t/d of feed lignite are provided.

.2 Section 4323 Shift Conversion

A shift conversion unit is provided to shift 28 percent of the syngas from the gasification section.

.3 Section 4350 Electrolysis Plant

An electrolysis plant with a total power consumption of 252 MW is required. Steam driven compressors are provided to deliver the produced oxygen and hydrogen to the gasifiers and methanol synthesis plant respectively.

.4 Section 4900 Utilities

Area 4913 Steam Generation

A lignite fuel gas fired steam boiler is provided to generate 5353 t/d of high pressure superheated steam to balance the steam system.

The high pressure steam is used for the steam turbine drives in the air separation and electrolysis plants, raw gas compression and methanol synthesis units. The medium pressure and low pressure steam balance description is generally as described for the base case and option 3 for the other technologies.

3.5.6 Thermal Efficiency

The overall thermal efficiencies for the base case and options are shown in Table 3-8.

TABLE 3-4
KOPPERS-TOTZEK UTILITY REQUIREMENTS

Utility	Base Case	Option[1] 1	Option[2] 2	Option[3] 3
Steam Generation[4], t/d	10900	2329	4136	8475
Cooling Water Circulation, t/d	345000	181000	192000	192000
Boiler Feed Water, t/d	2200	3800	1870	6330
Raw Water, t/d	11500	10600	8900	13140
Power, kW	9220	10800	14000	205400

Notes

1. Option 1 designates the natural gas co-processing option.
2. Option 2 designates the natural gas PSA co-processing option.
3. Option 3 designates the H_2 electrolysis option.
4. Steam generation designates offsite steam generation requirements in excess of onsite waste heat recovery systems. Steam conditions are 10.8 MPa and 480°C.

TABLE 3-5
LURGI UTILITY REQUIREMENTS

Utility	Base Case	Option A
Steam Generation[2], t/d	8034	7895
Cooling Water Circulation, t/d	240000	169000
Boiler Feed Water, t/d	5600	4700
Raw Water, t/d	11520	11620
Power, kW	20600	22170

Notes

1. Option A designates the conversion of methanol purge gas to syngas (H_2 + CO) via partial oxidation.

2. Steam generation designates offsite steam generation requirements in excess of onsite waste heat recovery systems. Steam conditions are 10.8 MPa and 480°C.

TABLE 3-6
WESTINGHOUSE UTILITY REQUIREMENTS

Utility	Base Case	Option[1] 1	Option[2] 2	Option[3] 3	Option[4] 4
Steam Generation[5], t/d	4000	1687	2636	2891	900
Steam Conditions	10.8 MPa @ 480°C	3.5 MPa	3.5 MPa	10.8 MPa @ 480°C	10.8 MPa @ 480°C
Cooling Water Circulation, t/d	281800	225000	234000	217000	218800
Boiler Feed Water, t/d	1650	1800	2570	1600	3100
Raw Water, t/d	11400	9740	10630	11730	15210
Power, kW	15400	9320	11740	223000	Nil

Notes

1. Option 1 designates the natural gas co-processing option.
2. Option 2 designates the natural gas PSA co-processing option.
3. Option 3 designates the H_2 electrolysis option.
4. Option 4 designates the H_2 electrolysis option with onsite co-generation of electricity.
5. Steam generation designates offsite steam generation requirements in excess of onsite waste heat recovery systems.

TABLE 3-7
HIGH TEMPERATURE WINKLER UTILITY REQUIREMENTS

Utility	Base Case	Option[1] 1	Option[2] 2	Option[3] 3
Steam Generation[4], t/d	7629	2445	4504	5353
Cooling Water Circulation, t/d	283000	275000	335000	217000
Boiler Feed Water, t/d	1660	1370	1900	1080
Raw Water, t/d	14700	11210	13900	8850
Power, kW	14700	11000	15000	263000

Notes

1. Option 1 designates the natural gas co-processing option.
2. Option 2 designates the natural gas PSA co-processing option.
3. Option 3 designates the H_2 electrolysis option.
4. Steam generation designates offsite steam generation requirements in excess of on site waste heat recovery systems. Steam conditions are 10800 kPa and 480°C.

TABLE 3-8

OVERALL THERMAL EFFICIENCY

		(Percent)
Koppers-Totzek	Base Case	42.0
	Option 1	63.9
	Option 2	57.9
	Option 3	46.0
Lurgi	Base Case	53.3
	Option A	50.7
Westinghouse	Base Case	51.1
	Option 1	65.7
	Option 2	59.9
	Option 3	57.9
	Option 4	43.0
H.T. Winkler	Base Case	48.7
	Option 1	65.2
	Option 2	58.1
	Option 3	52.8

Overall Thermal Efficiency is defined as follows:

Base Case:

$$\frac{\text{(Methanol.HHV)}}{\text{(Lignite to Gasifiers and Boiler.HHV) + Power*}}$$

Options 1 & 2:

$$\frac{\text{(Methanol HHV)}}{\text{(Lignite to Gasifiers \& Boiler.HHV) + (Natural Gas to Reformer Process and Fuel.HHV) + Power*}}$$

Option 3:

$$\frac{\text{(Methanol.HHV)}}{\text{(Lignite to Gasifiers and Boiler.HHV) + Power*}}$$

Option 4:

$$\frac{\text{(Methanol HHV)}}{\text{(Lignite to Gasifiers \& Combined Cycle Power Generation System.HHV)}}$$

* Power energy value taken at 100 percent efficiency (3.6 GJ/MWh)

Capital and Operating Costs

4.0 CAPITAL AND OPERATING COSTS

4.1 INTRODUCTION

Preliminary capital, operating and production cost estimates have been prepared for the base case and options of an Onakawana lignite-to-methanol plant. The capital and operating costs have been prepared in first quarter 1983 Canadian dollars for a Northern Ontario location. Operating costs include raw materials, utilities, manpower and maintenance requirements, municipal taxes and insurance. Product costs (F.O.B. plant) have been prepared taking into account operating costs, by-product credits and capital recovery factors. The sensitivity of production cost to capital recovery factors of 10, 15 and 20 percent is presented.

4.2 CAPITAL COST CRITERIA

The primary intent of Phase 1 is to carry out a financial comparison of the four competing technologies on a consistent basis. It is also the intent to generate total estimated capital and operating costs for a grass-roots lignite-to-methanol plant in Onakawana, Ontario within the framework of the Phase 1 scope and budget. As the proposed plant site is located in Onakawana, Ontario, a remote location accessible only by rail, there are many costs not included within the estimate which will require particular attention in future financial assessments. These include:

- Rail upgrading to allow shipment of heavy process and construction equipment.
- Road and rail construction outside plant battery limits.
- On-site concrete mix plant and backfill import requirements.
- Methanol distribution system costs such as tank cars, terminals and loading facilities.

4.2.1 Assumptions

The following basic assumptions were made for the financial assessment and comparison of the four gasification technologies: -

i) The plant is a grassroots facility, located at the lignite mine-mouth in Onakawana, Ontario. Lignite is delivered to raw material storage by conveyor. Rail transportation and rail off-loading facilities have not been considered in the estimate.

ii) Electricity is available from the existing grid in the local area to supply up to 50 MW of plant power consumption. This satisfies the purchased power requirements for the base case and Option 1, 2 and A evaluations. The electrolysis case (Option 3) for the Koppers-Totzek, Westinghouse and High Temperature Winkler processes entail a maximum load of 263 MW to be purchased from Ontario Hydro. This would require installation of a new power line approximately 48 km long at a cost of $195 000/km. The power line would tie into the existing grid from the Otter Rapids generating station and would be routed north to the Onakawana site. The cost of this installation would be to the project. Ontario Hydro verbally confirmed this information.

iii) Raw water to the plant is available from a river adjacent to the plant at no cost to the Owner.

iv) The co-processing options of natural gas and lignite entail construction of a new natural gas pipeline to the plant gate from the Trans Canada Pipeline. Inquiries to Northern and Central Gas, the operator in the area confirmed that a 323.8 mm dia. pipeline approximately 180 km long would be required to supply 850 000 Nm³/d. The line would tie into the existing Trans Canada Pipeline at Smooth Rock Falls, Ontario and would be installed parallel to the existing highway and railroad northward to Onakawana. The installation cost of the pipeline would not be capitalized, but rather included as a demand charge with the operating rate. Thus the installation cost is reflected in the operating costs.

v) Due to the remoteness of the location, a construction camp is provided to house construction and project personnel throughout the construction period. This cost is estimated to be 9.5 percent of the total direct cost of the plant.

vi) Operating labour costs are based on the average rate predominant in Northern Ontario mining locations. No allowance has been made for townsite or travelling costs.

vii) Normal construction techniques for civil work such as excavation, backfill, foundations by spread footing construction etc. has been assumed. No allowance has been made for importation of backfill and special foundation construction such as piling.

4.2.2 Capital Costs

Capital costs were derived using Kilborn in-house estimating information and data from process developers and licensors. Where cost data were not specific to a Northern Ontario location, they were converted to a cost in U.S. dollars, Gulf Coast location. This cost was then converted from U.S. Gulf Coast to a Northern Ontario location using a factor Canadian Dollars, Northern Ontario location, = 1.5 x U.S. dollars, Gulf Coast location.

This conversion factor is made up of two components - a materials component representing 60% of the value of the plant and a labour component representing 40% of the plant cost.

For the materials component a factor of 1.07 is used to represent 18% duty on the imported material (about 40% of total material). An exchange factor of 1.25 is used to convert U.S. dollars to Canadian, and a factor of 1.05 is used to represent 5% freight on all material. This results in a materials factor of:

$$0.6\ (1.07 \times 1.25 \times 1.05) = 0.843$$

The labour factor is comprised of the ratio of the base labour rate for a representative trade, such as pipefitting, in Canada to that in the Gulf Cost, each expressed in its own currency. This is multiplied by a productivity factor for Northern Ontario compared to the Gulf Coast which is estimated to be 1.2. The labour factor is then:

$$0.4 \quad \frac{18.00}{14.81} \times 1.2 \qquad = 0.583$$

A winterization factor of 1.05 is applied to the combined total to reflect the northern location, necessitating extra insulation and heat tracing on plant piping, equipment and instrumentation, to give a total conversion factor:

$$(0.843 + 0.583)\ 1.05 = 1.497, \text{ use } 1.5.$$

Buildings and site development costs were factored from direct costs. Building costs were assessed as two percent of the process unit costs and site development as five percent of the total direct costs.

Costs are presented for each technology base case and options in Canadian dollars, Northern Ontario location. The costs are subtotalled for on-site process units and offsites, utilities and services. Site development and building costs then are added resulting in Total Direct Costs for the plant. A total of 20 percent of direct costs is added to represent home office engineering and indirect field costs such as tools, equipment rental and contractors fees. Contingency is applied to the total direct costs by taking into account complexity and the degree of confidence in the direct cost estimate for the individual process units.

The Koppers-Totzek, Westinghouse and Lurgi processes have a contingency of 16%, while the High Temperature Winkler has 14%. Total Installed Cost is computed by adding the Total Direct Costs to the Indirect Cost and Contingency figures.

As discussed previously a new power transmission line is included in Option 3 for the Koppers-Totzek, Westinghouse and High Temperature Winkler technologies. This is assessed at $10.8 million and applied accordingly.

Owners costs, which include Owner's engineering and project management costs, and an Initial Catalyst and Chemical Cost and the Construction Camp cost are added to the Total Installed Cost to give a Total Estimated Cost.

4.2.3 Operating Costs

Operating costs consist of the costs of raw materials, purchased utilities and services, labour, maintenance and overheads. Annual operating costs are derived consistent with the operating factors presented in Table 3.3.

Operating costs comprise the following:

.1 Raw Materials

Lignite

Lignite is assumed to be received at the plant gate, minus 50 mm size and containing 45% moisture. The cost of lignite supplied to the plant is $15.00/t. This cost has been confirmed by Manalta Coal as being the minimum acceptable cost.

Natural Gas

A cost of $0.165/Nm^3 has been provided by Northern and Central Gas Corp. as representative for the study. A pipeline demand charge of $8 236 000 per year to cover pipeline construction maintenance and operating costs is also levied. This charge is based on an annual service of 265 000 000 Nm^3, and is factored to reflect the different natural gas loads for the optional cases considered.

.2 Labour

The labour component comprises the following categories:

a) Operating Labour. The manpower content ranges from a low of 105 for the Koppers-Totzek natural gas Options 1 and 2 to a high of 144 for the Westinghouse co-generation Option 4. Labour cost is averaged at $30,000/yr.

b) Plant overhead costs, which are estimated as 30% of operating labour.

c) Local administration, to cover plant management and supervision, is estimated at 20% of operating labour.

d) Corporate overhead costs, which include head office and sales overhead allocation, are estimated at 10% of operating labour.

e) Burden and supplies, consisting of payroll burdens, office equipment, supplies etc., are estimated at 30% of operating labour.

.3 Insurance

Insurance is estimated at 0.5% of Total Installed Cost of the facilities.

.4 Taxes

Taxes are estimated at 0.25% of Total Installed Cost and represent municipal taxes applicable to the plant facilities.

.5 Maintenance

Maintenance costs are split into two categories, labour and materials as follows:

Labour: 1.5 x Operating Labour

Materials (Spare Parts, etc.):	1% of Total Installed Cost for Base Case Options 1, 2 & A.
:	1% of (Total Installed Cost minus cost of Electrolysis facilities) for Options 3 & 4 to reflect low maintenance electrolysis plants.

.6 Catalyst and Chemicals

2% of Process Unit Direct Costs.

.7 Utilities

Power - Ontario Hydro supplied a rate of 2.71 cents per kWh which applies for consumptions up to 500 MW.

Lignite - Lignite to the steam boilers is costed at the same rate as run-of-mine lignite, $15.00/t.

4.2.4 Production Costs

This cost is computed by subtracting the by-product credits from the Total Operating Cost.

.1 By-product Credits

Saleable by-products are only considered for credit. In the Lurgi process, tars, tar oils and phenols are produced but are considered as having no market in Northern Ontario; hence these products are burnt in the steam boiler as fuel. Solid sulphur, anhydrous ammonia and lignite fines have been assessed the following credits:

Sulphur:	\$100/t
Anhydrous Ammonia:	\$240/t
Lignite Fines:	\$ 0/t

Lignite fines generated in the lignite preparation area for the Lurgi gasification technology have been estimated at 25 percent of the total lignite feed to the plant. About 9.5 percent of the fines are utilized in the steam boiler and the excess are assumed to be sold at zero value (plant gate) to an industrial user, such as the Ontario Hydro lignite fired generating station in Thunder Bay, or disposed of as solid waste.

4.2.5 Total Product Cost

Total Product Cost has been presented for three Capital Recovery Factors, 10 percent/yr, 15 percent/yr and 20 percent/yr. The Capital Recovery Cost in \$/tonne is computed as follows:

$$\frac{\text{Total Estimated Costs (\$)}}{\text{Capacity of Plant (tonne/yr)}} \times \frac{\text{Percent Capital Recovery}}{\text{Yr}}$$

The total of the Production Cost and the Capital Recovery Cost equals the Total Product Cost.

4.3 ESTIMATES

The Capital and Operating Costs are presented in Canadian dollars on the following Tables:

Technology	Table Number	
	Total Direct Cost	Operating Cost
Koppers-Totzek	4.1	4.6
Lurgi	4.2	4.7
Westinghouse	4.3	4.8
High Temperature Winkler	4.4	4.9

A summary of Capital Costs in Canadian dollars, encompassing the base cases and options for all four technologies, is shown on Table 4.5.

A summary of Capital Costs in Canadian dollars, Operating and Production Costs in Canadian dollars per tonne methanol is given in Table 4.10. The Product Cost in Canadian dollars per tonne methanol is also given at 10 percent, 15 percent and 20 percent Capital Recovery Factors.

TABLE 4-1

DIRECT COST - KOPPERS TOTZEK TECHNOLOGY

Estimated Costs in Million Canadian Dollars, 1st Quarter 1983.

SECTION	AREA	DESCRIPTION	BASE CASE	OPTION 1	OPTION 2	OPTION 3
1100		LIGNITE PREPARATION				
	1110	Size Reduction	9.3	4.0	6.1	6.8
	1120	Drying	25.0	10.9	16.5	18.4
1200		GASIFICATION				
	1210	Gasifiers				
	1220	Waste Heat Boilers	83.9	38.5	38.5	54.5
	1230	Gas Quench				
1300		GAS COMPRESSION & TREATING				
	1310	Syngas Compression	32.0	13.9	21.1	26.2
	1320	CO Shift Conversion	13.4	----	----	10.6
	1330	Acid Gas Removal	48.6	13.4	8.4	36.7
1400		METHANOL SYNTHESIS & PURIFICATION				
	1410	Methanol Synthesis	47.8	56.4	55.7	47.8
	1420	Methanol Purification	12.0	14.4	13.2	12.0
1500		AIR SEPARATION, GAS CONVERSION, ELECTROLYSIS				
	1510	Air Separation	55.2	24.0	36.4	31.2
	1520	Steam Methane Reforming	----	71.5	71.0*	----
	1530	Partial Oxidation	----	----	----	----
	1540	Pressure Swing Adsorption	----	----	32.7	----
	1550	Electrolysis Plant	----	----	----	59.0**
1600		SULPHUR PROCESSING				
	1610	Sulphur Recovery	5.9	2.6	3.9	5.1
	1620	Sulphur Slating	1.7	0.8	1.2	1.5
	1630	Tail Gas Processing	4.0	1.7	2.6	3.5
1700		GAS LIQUOR SEPARATION	----	----	----	----
		SUB TOTAL PROCESS UNITS	338.8	252.1	307.3	313.3
1800		EFFLUENT TREATMENT & DISPOSAL				
	1810	Fines Settling, Ash Dewatering	Incl in 1200	Incl in 1200	Incl in 1200	Incl in 1200
	1820	Solids Disposal	3.4	1.5	2.2	2.9
	1830	Phenosolvan, Ammonia Recovery	----	----	----	----
	1840	Wastewater Treatment	6.4	6.4	6.4	6.4
1900		UTILITIES, BUILDINGS & OFFSITES				
	1910	Steam Generation	29.7	11.5	18.3	26.1
	1920	Power Distribution	5.5	2.9	4.1	11.2
	1930	Cooling Water System	3.4	2.6	2.8	2.8
	1940	Raw Water System	3.3	3.1	3.0	3.0
	1950	Boiler Feed Water System & Demineralized Water	13.2	18.6	12.2	8.6
	1960	Firewater	1.4	1.4	1.4	1.4
	1970	Buildings	6.8	5.0	6.1	6.2
	1980	Offsites	32.2	37.4	36.5	32.2
		SUB TOTAL UTILITIES & SERVICES	105.3	90.4	93.0	100.8
		TOTAL	444.1	342.5	400.3	414.1
		SITE DEVELOPMENT @ 5%	22.2	17.1	20.8	20.7
		TOTAL DIRECT COST	466.3	359.6	420.3	434.8

* Includes H.T. Shift conversion.

** Includes H_2 and O_2 compression.

TABLE 4-2

DIRECT COST - LURGI GASIFICATION TECHNOLOGY

Estimated Costs in Million Canadian Dollars, 1st Quarter 1983.

SECTION	AREA	DESCRIPTION	BASE CASE	OPTION A
2100		LIGNITE PREPARATION		
	2110	Size Reduction	6.5	6.5
	2120	Drying	12.0	12.0
2200		GASIFICATION		
	2210	Gasifiers		
	2220	Waste Heat Boilers	63.0	63.0
	2230	Gas Quench	3.7	3.7
2300		GAS COMPRESSION & TREATING		
	2310	Syngas Compression	----	----
	2320	CO Shift Conversion	----	----
	2330	Acid Gas Removal	45.0	45.0
2400		METHANOL SYNTHESIS & PURIFICATION		
	2410	Methanol Synthesis	49.9	49.9
	2420	Methanol Purification	12.1	12.1
2500		AIR SEPARATION, GAS CONVERSION, ELECTROLYSIS		
	2510	Air Separation	29.1	36.6
	2520	Steam Methane Reforming	37.7	----
	2530	Partial Oxidation	----	37.5
	2540	Pressure Swing Adsorption	----	----
	2550	Electrolysis Plant	----	----
2600		SULPHUR PROCESSING		
	2610	Sulphur Recovery	6.6	6.6
	2620	Sulphur Slating	1.8	1.8
	2630	Tail Gas Processing	4.4	4.4
2700		GAS LIQUOR SEPARATION	6.5	6.5
		SUB TOTAL PROCESS UNITS	278.3	285.6
2800		EFFLUENT TREATMENT & DISPOSAL		
	2810	Fines Settling, Ash Dewatering	8.7	8.7
	2820	Solids Disposal	7.9	7.9
	2830	Phenosulvan, Ammonia Recovery	15.9	15.9
	2840	Wastewater Treatment	18.0	18.0
2900		UTILITIES, BUILDINGS & OFFSITES		
	2910	Steam Generation	27.6	27.2
	2920	Power Generation	6.9	7.2
	2930	Cooling Water System	3.4	3.7
	2940	Raw Water System	3.4	3.2
	2950	Boiler Feed Water System	22.9	22.6
	2960	Firewater	1.4	1.4
	2970	Buildings	5.6	5.7
	2980	Offsites	36.6	36.6
		SUB TOTAL UTILITIES & SERVICES	159.3	158.1
		TOTAL	437.6	443.7
		SITE DEVELOPMENT @ 5%	21.9	22.2
		TOTAL DIRECT COST	459.5	465.9

TABLE 4-3

DIRECT COST - WESTINGHOUSE

Estimated Costs in Million Canadian Dollars, 1st Quarter 1983.

SECTION	AREA	DESCRIPTION	BASE CASE	OPTION 1	OPTION 2	OPTION 3	OPTION 4
3100		LIGNITE PREPARATION					
	3110	Size Reduction	5.7	2.5	3.8	4.7	6.8
	3120	Drying	15.9	6.9	10.5	13.1	19.0
3200		GASIFICATION					
	3210	Gasifiers					
	3220	Waste Heat Boilers	85.8	42.9	42.9	42.9	42.9
	3230	Gas Quench					
3300		GAS COMPRESSION & TREATING					
	3310	Syngas Compression	----	----	----	----	----
	3320	CO Shift Conversion	14.3	----	----	9.9	9.9
	3330	Acid Gas Removal	43.7	14.1	10.2	32.2	32.0
3400		METHANOL SYNTHESIS & PURIFICATION					
	3410	Methanol Synthesis	47.5	50.6	50.1	47.5	47.5
	3420	Methanol Purification	11.8	13.2	12.9	11.8	11.8
3500		AIR SEPARATION, GAS CONVERSION, ELECTROLYSIS					
	3510	Air Separation	48.4	21.1	32.0	16.7	16.7
	3520	Steam Methane Reforming	19.1	60.4	62.9*	----	----
	3530	Partial Oxidation	----	----	----	----	----
	3540	Pressure Swing Adsorption	----	----	24.5	----	----
	3550	Electrolysis Plant	----	----	----	80.5**	80.5**
3600		SULPHUR PROCESSING					
	3610	Sulphur Recovery	6.6	2.9	4.4	4.4	4.4
	3620	Sulphur Slating	1.8	0.8	1.2	1.2	1.2
	3630	Tail Gas Processing	4.4	1.9	2.9	2.9	2.9
3700		GAS LIQUOR SEPARATION	----	----	----	----	----
		SUB TOTAL PROCESS UNITS	305.0	217.3	258.3	267.8	275.6
3800		EFFLUENT TREATMENT & DISPOSAL					
	3810	Fines Settling, Ash Dewatering	5.5	2.4	4.6	4.5	6.5
	3820	Solids Disposal	4.1	1.8	2.7	2.7	2.7
	3830	Phenosolvan, Ammonia Recovery	2.7	1.2	1.8	1.7	2.5
	3840	Wastewater Treatment	6.4	6.4	6.4	6.4	6.4
3900		UTILITIES, BUILDINGS & OFFSITES					
	3910	Steam Generation	18.0	6.7	9.9	15.3	6.1
	3920	Power Generation & Distribution	7.0	3.4	4.2	11.2	222.9
	3930	Cooling Water System	4.0	3.2	3.3	3.2	3.4
	3940	Raw Water System	4.0	3.0	3.1	3.4	5.4
	3950	Boiler Feed Water System & Demineralized Water	12.7	6.9	8.7	14.5	18.7
	3960	Firewater	1.4	1.4	1.4	1.4	1.4
	3970	Buildings	6.1	4.3	5.2	5.6	6.2
	3980	Offsites	33.8	32.3	35.1	33.8	30.3
		SUB TOTAL UTILITIES & SERVICES	105.7	73.0	86.4	103.7	312.5
		TOTAL	410.7	290.3	344.7	371.5	588.1
		SITE DEVELOPMENT @ 5%	20.5	14.5	17.2	18.6	29.4
		TOTAL DIRECT COST	431.2	304.8	361.9	390.1	617.5

* Includes H.T. shift conversion.

** Includes H_2 and O_2 compression.

TABLE 4-4

DIRECT COST - HIGH TEMPERATURE WINKLER

Estimated Costs in Million Canadian Dollars, 1st Quarter 1983.

SECTION	AREA	DESCRIPTION	BASE CASE	OPTION 1	OPTION 2	OPTION 3
4100		LIGNITE PREPARATION				
	4110	Size Reduction	3.8	2.0	3.0	3.0
	4120	Drying	18.2	9.6	14.6	14.6
4200		GASIFICATION				
	4210	Gasifiers				
	4220	Waste Heat Boilers	132.2	66.1	66.1	66.1
	4230	Gas Quench				
4300		GAS COMPRESSION & TREATING				
	4310	Syngas Compression	17.1	6.4	9.7	9.7
	4320	CO Shift Conversion	13.5	----	----	8.0
	4330	Acid Gas Removal	41.7	17.3	15.1	26.8
4400		METHANOL SYNTHESIS & PURIFICATION				
	4410	Methanol Synthesis	50.0	57.7	63.9	50.1
	4420	Methanol Purification	12.3	16.4	16.8	12.3
4500		AIR SEPARATION, GAS CONVERSION, ELECTROLYSIS				
	4510	Air Separation	51.1	26.4	40.0	21.9
	4520	Steam Methane Reforming	----	66.6	61.1*	----
	4530	Partial Oxidation	----	----	----	----
	4540	Pressure Swing Adsorption	----	----	41.2	----
	4550	Electrolysis Plant	----	----	----	87.2**
4600		SULPHUR PROCESSING				
	4610	Sulphur Recovery	6.0	3.1	4.8	4.8
	4620	Sulphur Slating	1.7	0.9	1.3	1.3
	4630	Tail Gas Processing	4.1	2.1	3.2	3.2
4700		GAS LIQUOR SEPARATION	----	----	----	----
		SUB TOTAL PROCESS UNITS	351.9	274.6	340.8	304.0
4800		EFFLUENT TREATMENT & DISPOSAL				
	4810	Fines Settling, Ash Dewatering	Incl in 4200	Incl in 4200	Incl in 4200	Incl in 4200
	4820	Solids Disposal	2.6	1.3	2.0	2.0
	4830	Phenosulvan, Ammonia Recovery	3.8	2.0	3.0	2.4
	4840	Wastewater Treatment	6.4	6.4	6.4	6.4
4900		UTILITIES, BUILDINGS & OFFSITES				
	4910	Steam Generation	23.7	11.2	17.5	20.8
	4920	Power Distribution	5.6	3.9	5.1	4.5
	4930	Cooling Water System	3.9	3.8	4.6	3.1
	4940	Raw Water System	3.4	3.3	3.7	3.1
	4950	Boiler Feed Water System & Demineralized Water	11.4	7.2	9.2	10.3
	4960	Firewater	1.4	1.4	1.4	1.4
	4970	Buildings	6.8	5.6	6.9	6.3
	4980	Offsites	33.7	36.8	37.6	33.7
		SUB TOTAL UTILITIES & SERVICES	102.7	83.9	97.4	94.0
		TOTAL	454.6	358.5	438.2	403.0
		SITE DEVELOPMENT @ 5%	22.7	17.9	21.9	20.1
		TOTAL DIRECT COST	477.3	376.4	460.1	423.1

* Includes H.T. shift conversion.

** Includes H_2 and O_2 compression.

TABLE 4-5

SUMMARY OF TOTAL ESTIMATED COSTS - ALL TECHNOLOGIES

(Million Canadian Dollars)

UNIT NO.	DESCRIPTION	KOPPERS-TOTZEK				LURGI		WESTINGHOUSE					HIGH TEMPERATURE WINKLER			
		BASE CASE	OPTION 1	OPTION 2	OPTION 3	BASE CASE	OPTION A	BASE CASE	OPTION 1	OPTION 2	OPTION 3	OPTION 4	BASE CASE	OPTION 1	OPTION 2	OPTION 3
100	Lignite Preparation	34.3	14.9	22.6	25.2	18.5	18.5	21.6	9.4	14.3	17.8	25.8	22.0	11.6	17.6	17.6
200	Gasification	83.9	38.5	38.5	54.5	66.7	66.7	85.8	42.9	42.9	42.9	42.9	132.2	66.1	66.1	66.1
300	Gas Compression & Treating	94.0	27.3	29.5	73.5	45.0	45.0	58.0	14.1	10.2	42.1	42.1	72.3	23.7	24.8	44.5
400	Methanol Synthesis & Purification	59.8	70.8	68.9	59.8	62.0	62.0	59.3	63.8	63.0	59.3	59.3	62.3	74.1	80.7	62.4
500	Air Separation, SMR, PSA, Catalytic Oxidation, Electrolysis	55.2	95.5	140.1	90.2	66.8	74.1	67.5	81.5	119.4	97.2	97.2	51.1	93.0	142.3	109.1
600	Sulphur Processing	11.6	5.1	7.7	10.1	12.8	12.8	12.8	5.6	8.5	8.5	8.5	11.8	6.1	9.3	9.3
700	Gas Liquor Separation	----	----	----	----	6.5	6.5	----	----	----	----	----	----	----	----	----
	SUB-TOTAL PROCESS UNITS	338.8	252.1	307.3	313.3	278.3	285.6	305.0	217.3	258.3	267.8	275.6	351.9	274.6	340.8	309.0
800	Effluent Treatment & Disposal	9.8	7.9	8.6	9.3	50.5	50.5	18.7	11.8	15.5	15.3	18.1	12.8	9.7	11.4	10.8
900	Utilities, Buildings & Offsites	95.5	82.5	84.4	91.5	108.8	107.6	87.0	61.0	70.9	88.4	294.4	89.9	74.2	86.0	83.2
	SUB-TOTAL UTILITIES & SERVICES	105.3	90.4	93.0	100.8	159.3	158.1	105.7	73.0	86.4	103.7	312.5	102.7	83.9	97.4	94.0
	TOTAL	444.1	342.5	400.3	414.1	437.6	443.7	410.7	290.3	344.7	371.5	588.1	454.6	358.5	438.2	403.0
	Site Development @ 5%	22.2	17.1	20.0	20.7	21.9	22.2	20.5	14.5	17.2	18.6	29.4	22.7	17.9	21.9	20.1
	TOTAL DIRECT COST	466.3	359.6	420.3	434.8	459.5	465.9	431.2	304.8	361.9	390.1	617.5	477.3	376.4	460.1	423.1
	INDIRECT COSTS															
	Engineering & Construction Fees @ 20%	93.3	71.9	84.1	87.0	91.9	93.2	86.2	61.0	72.4	78.0	123.5	95.4	75.3	92.0	84.6
	Contingency @ 16%	74.6	57.5	67.2	69.6	73.5	74.5	69.0	48.8	57.9	62.4	98.8				
	@ 14%												66.8	52.7	64.4	59.2
	TOTAL INSTALLED COST	634.2	489.0	571.6	591.4	624.9	633.6	586.4	414.6	492.2	530.5	839.8	639.5	504.4	616.5	556.9
	Power Line for Option 3	----	----	----	10.8	----	----	----	----	----	10.8	----	----	----	----	10.8
	Owners Costs, Catalyst & Chemicals	12.7	9.8	11.4	11.8	12.5	12.7	11.7	8.3	9.8	10.6	16.8	12.8	10.1	12.3	11.3
	Construction Camp	44.3	34.2	39.9	41.3	43.7	44.3	41.0	29.0	34.4	37.1	58.7	45.3	35.7	43.7	40.2
	TOTAL ESTIMATED COST	691.2	533.0	622.9	655.3	681.1	690.6	639.1	451.9	536.4	589.0	915.3	697.6	550.2	672.5	629.2

TABLE 4-6
KOPPERS-TOTZEK ANNUAL OPERATING COST
(Million Canadian Dollars)

Item	Base Case	Option 1	Option 2	Option 3
Operating Factor (days)	320	320	317	313
Raw Materials				
- Lignite	28.3	7.1	14.0	20.7
- Natural Gas	----	79.4	64.0	----
Labour				
- Operating Labour	3.5	3.2	3.2	3.5
- Plant Overhead	1.0	1.0	1.0	1.0
- Local Administration	.7	.6	.6	.7
- Corporate Overhead	.3	.3	.3	.3
- Burden & Supplies	1.5	1.3	1.3	1.5
Insurance	3.4	2.5	2.8	3.0
Taxes	1.7	1.2	1.4	1.5
Maintenance	12.1	9.7	10.4	10.4
Catalyst & Chemicals	6.8	5.0	6.1	6.1
Utilities				
- Power	1.9	2.2	2.9	41.8
- Lignite	15.6	2.5	6.2	10.4
- Natural Gas	----	21.6	5.8	----
TOTAL	76.8	137.6	120.0	100.9

TABLE 4-7
LURGI ANNUAL OPERATING COST
(Million Canadian Dollars)

Item	Base Case	Option A
Operating Factor (days)	320	320
Raw Materials		
- Lignite	40.6	40.6
- Natural Gas	----	----
Labour		
- Operating Labour	3.8	4.1
- Plant Overhead	1.1	1.2
- Local Administration	.8	.8
- Corporate Overhead	.4	.4
- Burden & Supplies	1.5	1.7
Insurance	3.4	3.4
Taxes	1.7	1.7
Maintenance	12.5	13.1
Catalyst & Chemicals	5.6	5.7
Utilities		
- Power	4.3	4.6
- Lignite	----	----
TOTAL	75.7	77.3

TABLE 4-8
WESTINGHOUSE ANNUAL OPERATING COST
(Million Canadian Dollars)

Item	Base Case	Option 1	Option 2	Option 3	Option 4
Operating Factor (days)	320	320	317	313	313
Raw Materials					
- Lignite	28.2	7.4	14.6	20.9	20.9
- Natural Gas	----	59.7	51.5	----	----
Labour					
- Operating Labour	3.3	3.2	3.3	3.3	4.3
- Plant Overhead	1.0	.9	1.0	1.0	1.3
- Local Administration	.7	.7	.7	.7	.9
- Corporate Overhead	.3	.3	.3	.3	.4
- Burden & Supplies	1.3	1.3	1.3	1.3	1.7
Insurance	3.2	2.2	2.5	2.6	4.2
Taxes	1.6	1.1	1.3	1.3	2.1
Maintenance	11.3	9.3	10.3	9.7	14.5
Catalyst & Chemicals	6.1	4.3	5.2	5.5	5.5
Utilities					
- Power	3.2	1.9	2.4	45.3	----
- Lignite	6.0	2.5	3.2	1.7	21.0
- Natural Gas	----	17.6	----	----	----
TOTAL	66.2	112.4	97.6	93.6	76.8

TABLE 4-9
HIGH TEMPERATURE WINKLER ANNUAL OPERATING COST
(Million Canadian Dollars)

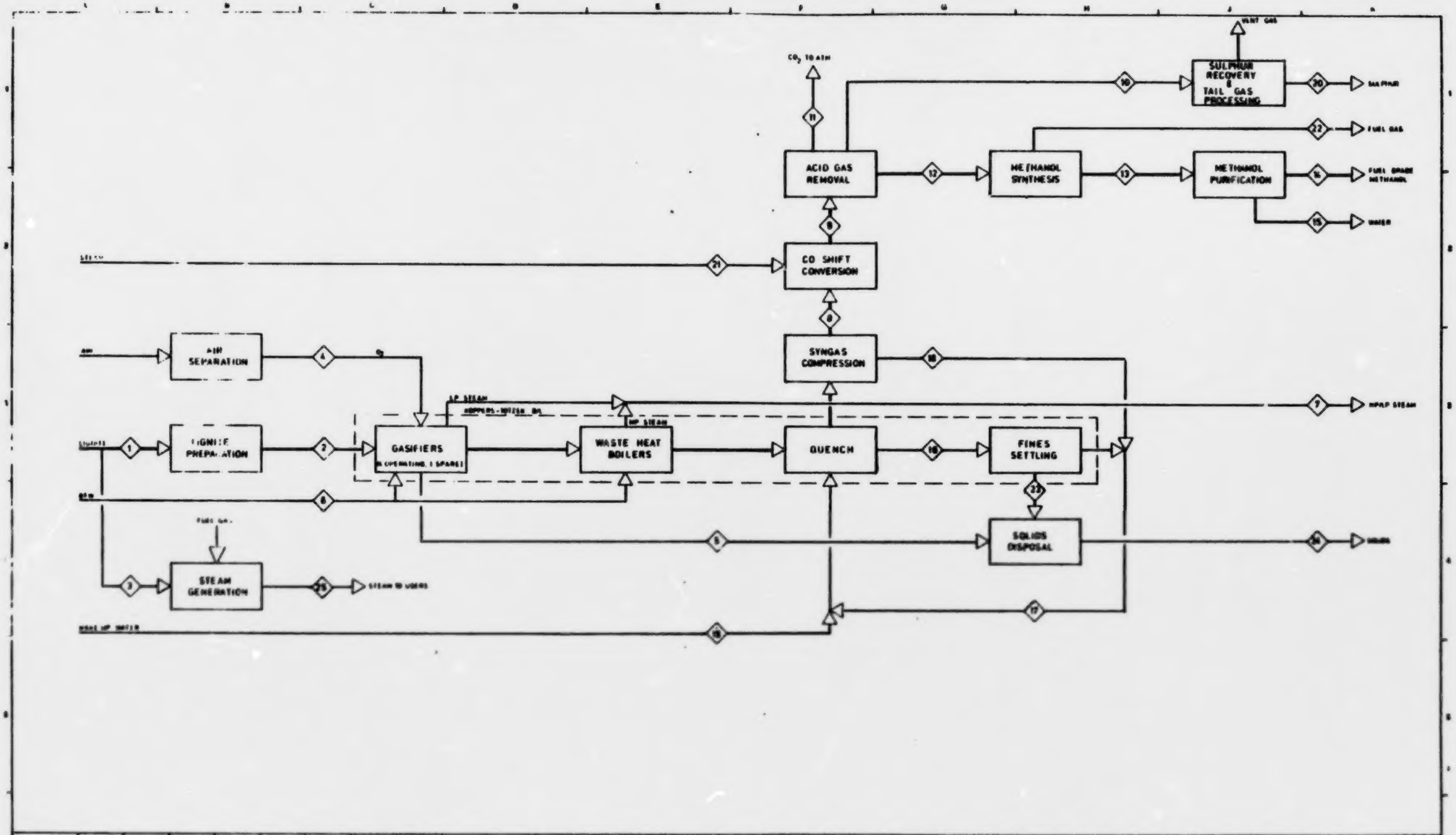
Item	Base Case	Option 1	Option 2	Option 3
Operating Factor (days)	320	320	317	313
Raw Materials				
- Lignite	26.9	9.2	18.2	18.0
- Natural Gas		71.1	61.1	
Labour				
- Operating Labour	3.4	3.4	3.4	3.4
- Plant Overhead	1.0	1.0	1.0	1.0
- Local Administration	.7	.7	.7	.7
- Corporate Overhead	.3	.3	.3	.3
- Burden & Supplies	1.4	1.4	1.4	1.4
Insurance	3.4	2.7	3.0	2.8
Taxes	1.7	1.8	1.5	1.4
Maintenance	12.0	10.6	11.2	10.2
Catalyst & Chemicals	7.0	5.5	6.9	6.4
Utilities				
- Power	3.1	2.3	3.1	53.5
- Lignite	9.3	3.7	6.8	6.3
- Natural Gas		24.6	6.5	
TOTAL	70.2	138.3	125.1	105.4

TABLE 4-19

SUMMARY OF CAPITAL AND OPERATING COST ESTIMATES

GASIFIER	KOPPERS-TOTZEK				LURGI		WESTINGHOUSE					HIGH TEMPERATURE WINKLER			
CASE	BASE CASE	OPTION 1	OPTION 2	OPTION 3	BASE CASE	OPTION A	BASE CASE	OPTION 1	OPTION 2	OPTION 3	OPTION 4	BASE CASE	OPTION 1	OPTION 2	OPTION 3
Operating Factor, days/year	320	320	317	313	320	320	320	320	317	313	313	320	320	317	313
Methanol Capacity, tonne/year	643200	847360	802960	629130	643200	643200	643200	712640	679650	629130	629130	643200	889600	898690	629130
CAPITAL COST, $ MM, CAN.															
Process Units	338.8	252.1	307.3	313.3	278.3	285.6	305.0	217.3	258.3	267.8	275.6	351.9	274.6	340.8	309.0
Other Services & Site Dev.	127.5	107.5	113.0	121.5	181.2	178.2	126.2	87.5	103.6	122.3	341.9	125.4	101.8	119.3	114.1
Engineering, Indirects, Contingency	167.9	129.4	151.3	156.6	165.4	167.0	155.2	109.8	130.3	140.4	222.3	162.2	128.0	156.4	143.8
Total Installed Cost	634.2	489.0	571.6	591.4	624.9	630.8	586.4	414.6	492.2	530.5	839.8	639.5	504.4	616.5	566.9
Owners Costs, Camp, Catalyst	57.0	44.0	51.3	63.9	56.2	56.7	52.7	37.3	44.2	58.5	75.5	58.1	45.8	56.0	62.3
TOTAL ESTIMATED COST $ MM	691.2	533.0	622.9	655.3	681.1	687.5	639.1	451.9	536.4	589.0	915.3	697.6	550.2	672.5	629.2
TOTAL ESTIMATED COST $/t METHANOL	1074.6	629.0	775.7	1041.6	1058.9	1068.9	993.6	634.1	789.2	936.2	1454.9	1084.6	618.5	748.3	1000.0
OPERATING COSTS, $/tonne															
Raw Materials - Lignite	44.00	8.38	17.44	32.90	63.12	63.12	43.85	10.38	21.48	33.22	33.22	41.82	10.34	20.25	28.61
- Natural Gas	----	93.70	79.70	----	----	----	----	83.77	75.77	----	----	----	79.92	67.99	----
Labour, Overhead, Admin, Supplies	10.88	7.55	7.97	11.13	11.82	12.75	10.26	8.98	9.71	10.49	13.67	10.57	7.64	7.57	10.81
Insurance & Municipal Taxes	7.93	4.37	5.23	7.15	7.92	7.92	7.46	4.63	5.59	6.20	10.01	7.93	5.06	5.01	6.67
Maintenance	18.81	11.45	12.95	16.53	19.43	20.37	17.56	13.05	15.15	15.42	23.05	18.65	11.91	12.46	16.21
Catalyst & Chemicals	10.57	5.90	7.60	9.70	8.71	8.86	9.48	6.03	7.65	8.74	8.74	10.88	6.18	7.68	10.17
Utilities - Power	2.95	2.60	3.61	66.44	6.69	7.15	4.98	2.66	3.53	72.00	----	4.82	2.58	3.45	85.04
- Lignite	24.25	2.95	7.72	16.53	----	----	9.33	3.51	4.71	2.70	33.34	14.46	4.15	7.51	10.01
- Natural Gas		25.49	7.22	-----				24.70	-----	-----	-----		27,65	7.23	
TOTAL OPERATING COST $/tonne	119.39	162.39	149.44	160.38	117.69	120.17	102.92	157.71	143.59	148.77	122.03	109.13	155.43	139.21	167.52
By-Product Credits															
- Sulphur @ $100/t	2.34	0.42	0.91	1.74	2.44	2.44	2.39	0.54	1.16	1.79	1.79	2.79	0.69	1.32	1.89
- Ammonia @ $240/t	----	----	----	----	4.18	4.18	.83	0.19	0.38	0.39	0.39	1.43	0.35	0.67	0.96
- Lignite Fines $0/t	----	----	----	----	0.00	0.00	----	----	----	----	----	----	----	----	----
Total By-Products Credits	2.34	0.42	0.91	1.74	6.62	6.62	3.22	0.73	1.54	2.18	2.18	4.22	1.04	1.99	2.85
TOTAL PRODUCTION COST, $/tonne (Operating Cost minus By-Product Credit)	116.75	161.97	148.53	158.64	111.07	113.55	99.70	156.98	142.05	146.59	119.85	104.91	154.39	137.22	164.67
TOTAL PRODUCT COST, $/tonne [Production Cost + (Total Est. Cost/t x percent Capital Recovery)]															
at 10% Capital Recovery	224.51	224.87	226.10	262.80	216.96	220.44	199.06	220.39	220.97	240.21	265.40	213.37	216.24	212.05	264.67
at 15% Capital Recovery	278.24	256.32	264.88	314.88	269.91	273.89	248.74	252.09	260.43	287.02	338.09	267.60	247.16	249.47	314.67
at 20% Capital Recovery	331.97	287.77	303.67	366.96	322.85	327.33	298.42	283.80	299.89	333.83	410.83	321.83	278.09	286.88	364.67

Kilborn has followed standard professional procedures in preparing the contents of this report, but it is based in part on details, information and assumptions provided by others and Kilborn, therefore, cannot guarantee the correctness of the report; but to the extent of its investigation and within the scope of the work delegated to it, Kilborn believes that the report is substantially correct.


AIR
AIR SEPARATION
O_2
LIGNITE
LIGNITE PREPARATION
BFW
FUEL GAS
STEAM GENERATION
STEAM TO USERS
MAKE-UP WATER
STEAM
GASIFIERS
(3 OPERATING, 1 SPARE)
LP STEAM
KOPPERS-TOTZEK BL
HP STEAM
WASTE HEAT BOILERS
QUENCH
FINES SETTLING
SOLIDS DISPOSAL
SYNGAS COMPRESSION
CO SHIFT CONVERSION
ACID GAS REMOVAL
CO_2 TO ATM
SULPHUR RECOVERY & TAIL GAS PROCESSING
VENT GAS
SULPHUR
METHANOL SYNTHESIS
METHANOL PURIFICATION
FUEL GAS
FUEL GRADE METHANOL
WATER
HP/LP STEAM
SOLIDS
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25

STREAM No	DESCRIPTION	FLOW RATE t/d
1	AS RECEIVED LIGNITE TO PREP (35% MOIST.)	5091
2	LIGNITE TO GASIFIERS ([illegible] MOIST.)	2522
3	LIGNITE TO STEAM PLANT (35% MOIST.)	3253
4	OXYGEN	2616
5	SLAG (DRY BASIS)	313
6	BFW TO WASTE HEAT BOILER	4163
7	HP STEAM ([illegible]) & LP STEAM ([illegible])	2150 1015
8	RAW SYNGAS	5279
9	SHIFTED SYNGAS TO AGR	6371
10	ACID GAS TO CLAUS PLANT	50
11	CO_2 TO ATM	3821
12	SYNGAS TO MeOH SYNTHESIS	3608
13	RAW METHANOL	2113
14	FUEL GRADE METHANOL	2090
15	WATER	103
16	WET FINES TO SETTLING	873
17	RECYCLE WATER TO QUENCH	865
18	WATER	345
19	MAKE-UP WATER	81
20	SULPHUR	47
21	STEAM TO CO SHIFT	2844
22	FUEL GAS	287
23	WET FINES TO DISPOSAL	353
24	SOLIDS TO DISPOSAL	666
25	HP SUPERH STEAM 480°C ([illegible] kPa)	10600
26		
27		
28		
29		
30		

REFERENCE DRAWINGS

No	DESCRIPTION	DATE	BY
C	GENERAL UPDATE	[illegible]	
B	RELEASED FOR FINAL REPORT	[illegible]	
A	RELEASED FOR DRAFT REPORT	[illegible]	

REVISIONS

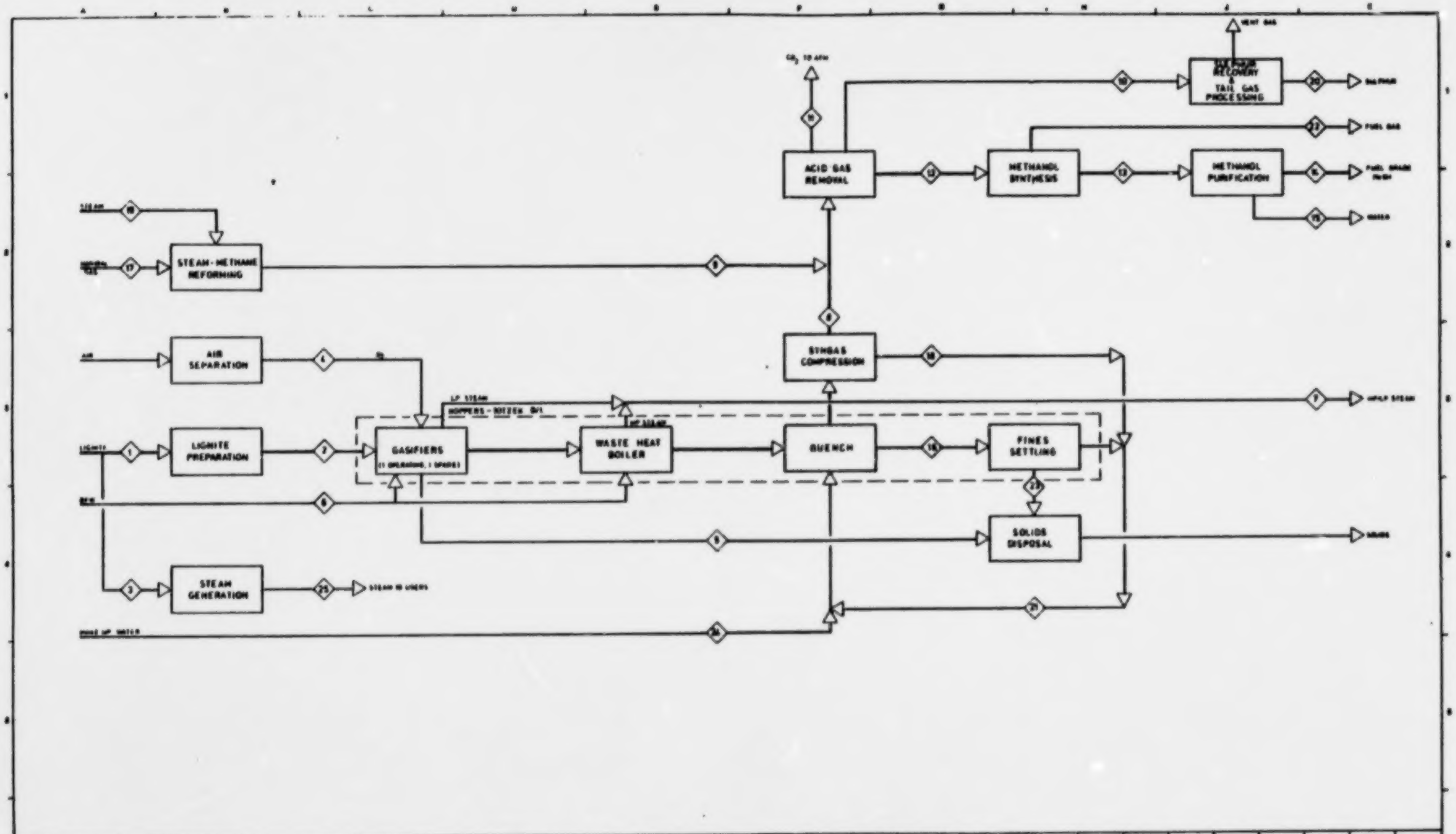
DESIGN PROCESS
SCALE NONE
DRAWN AS

ONTARIO ENERGY CORPORATION
LOCATION TORONTO ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
KOPPERS-TOTZEK
BASE CASE

3426
DRAWING NUMBER
100 43-01
REV C


STEAM
18
NATURAL GAS
17
STEAM-METHANE REFORMING
9
AIR
AIR SEPARATION
4
O_2
LIGNITE
1
LIGNITE PREPARATION
2
GASIFIERS (1 OPERATING, 1 SPARE)
LP STEAM
KOPPERS - TOTZEK B/L
HP STEAM
WASTE HEAT BOILER
BFW
6
5
3
STEAM GENERATION
25
STEAM TO USERS
MAKE UP WATER
24
CO_2 TO ATM
11
ACID GAS REMOVAL
10
8
SYNGAS COMPRESSION
12
QUENCH
16
FINES SETTLING
23
SOLIDS DISPOSAL
21
METHANOL SYNTHESIS
13
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
VENT GAS
20
SULPHUR
22
FUEL GAS
14
FUEL GRADE MeOH
15
WATER
7
HP/LP STEAM
SOLIDS

STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
DESCRIPTION	AS RECEIVED LIGNITE TO PREP (30% MOIST)	LIGNITE TO GASIFIER (10% MOIST)	LIGNITE TO STEAM PLANT (35% MOIST)	OXYGEN	SLAG (DRY BASIS)	BFW TO WASTE HEAT BOILER	HP STEAM [illegible] & LP STEAM [illegible]	RAW SYNGAS	REFORMED GAS TO AGR	ACID GAS TO CLAUS PLANT	CO_2 TO ATM	SYNGAS TO MeOH SYNTHESIS	RAW METHANOL	FUEL GRADE METHANOL	WATER
FLOW RATE	1473	881	730	656	78	1030	538 494	1238	228	17	448	3404	2762	2848	134

STREAM No	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
DESCRIPTION	FINES TO SETTLING	NATURAL GAS	WATER	M.P. STEAM	SULPHUR	RECYCLE WATER TO QUENCH	FUEL GAS	WET FINES TO DISPOSAL	MAKE-UP WATER	HP SUPERH STEAM 480° C [illegible]					
FLOW RATE	260	660	50	2786	11	135	222	172	28	2320					

REFERENCE DRAWINGS

No	DESCRIPTION	DATE	BY
C	GENERAL UPDATE	23.3.84	
B	RELEASED FOR FINAL REPORT	10.6.83	
A	RELEASED FOR DRAFT REPORT	8.7.83	

REVISIONS

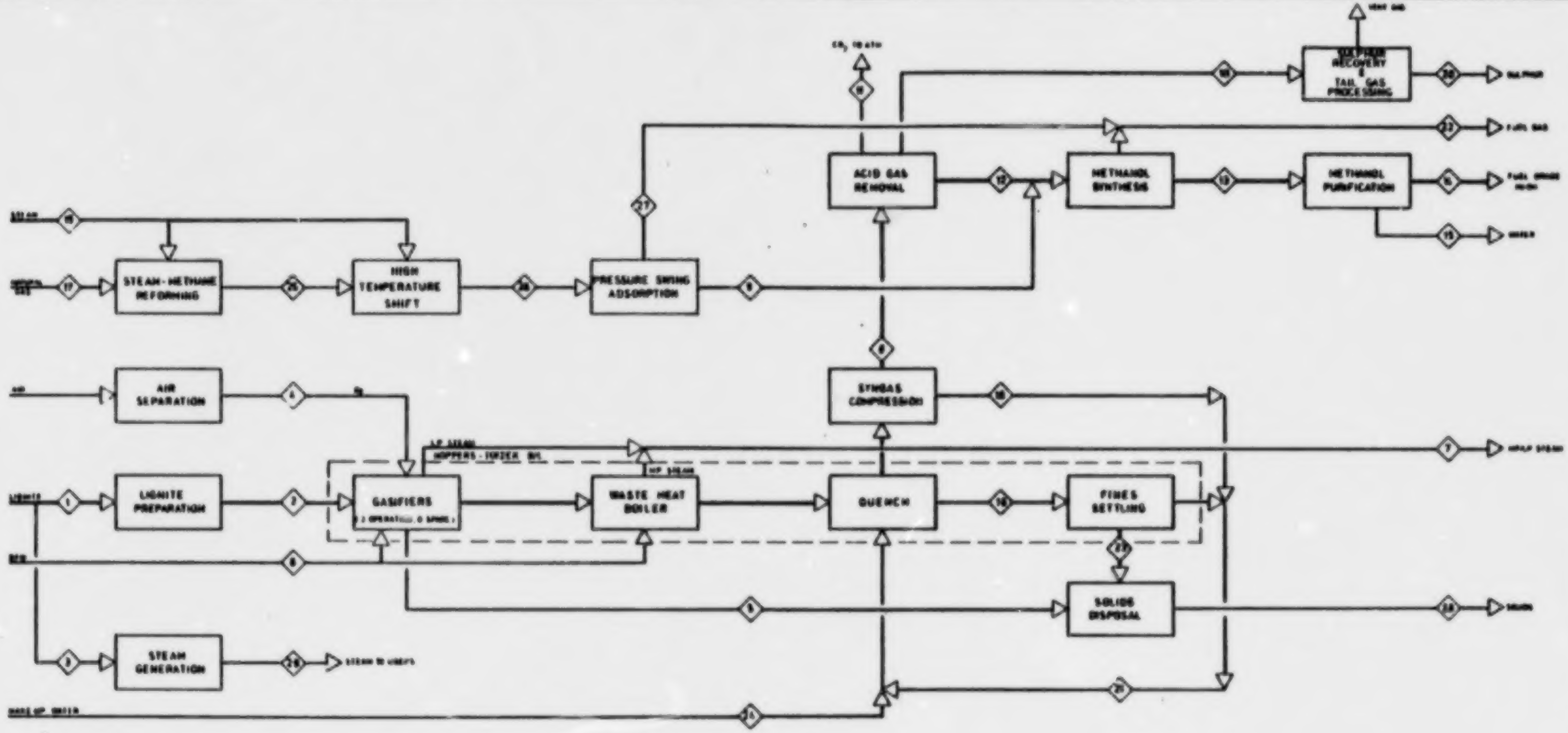
DESIGN PROCESS
SCALE NONE
DRAWN AS
CHECKED AS
APPROVED [illegible]

ONTARIO ENERGY CORPORATION
LOCATION TORONTO ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
KOPPERS - TOTZEK
OPTION 1
STEAM METHANE REFORMING

3474
18
DRAWING NUMBER
100-43-02
REV C


STEAM
NATURAL GAS
AIR
LIGNITE
BFW
MAKE UP WATER
STEAM-METHANE REFORMING
HIGH TEMPERATURE SHIFT
PRESSURE SWING ADSORPTION
AIR SEPARATION
O2
LIGNITE PREPARATION
GASIFIERS
(2 OPERATING, 0 SPARE)
LP STEAM
KOPPERS - TOTZEK B/L
HP STEAM
WASTE HEAT BOILER
STEAM GENERATION
STEAM TO USERS
ACID GAS REMOVAL
CO2 TO ATM
SYNGAS COMPRESSION
QUENCH
FINES SETTLING
SOLIDS DISPOSAL
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
VENT GAS
SULPHUR
FUEL GAS
FUEL GRADE MeOH
WATER
HP/LP STEAM
DRAIN
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
20
21
22
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STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
DESCRIPTION	AS RECEIVED LIGNITE TO PREP (35% MOIST.)	LIGNITE TO GASIFIERS (8% MOIST.)	LIGNITE TO STEAM PLANT (35% MOIST.)	OXYGEN	SLAG (DRY BASIS)	BFW TO WASTE HEAT BOILER	HP STEAM [illegible] & LP STEAM [illegible]	RAW SYNGAS	HYDROGEN	ACID GAS TO CLAUS PLANT	CO_2 TO ATM	SYNGAS TO MeOH SYNTHESIS	RAW METHANOL	FUEL GRADE METHANOL	WATER
FLOW RATE t/d	2946	1762	1298	1308	154	2082	1075 909	2675	307	25	0	2690	2687	2533	154

STREAM No	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
DESCRIPTION	FINES TO SETTLING	NATURAL GAS	WATER	STEAM	SULPHUR	RECYCLE WATER TO QUENCH	FUEL GAS	WET FINES TO DISPOSAL	MAKE UP WATER	SMR EFFLUENT	SHIFT EFFLUENT *	FUEL GAS	SOLIDS TO DISPOSAL	HP SUPERH. STEAM 480°C [illegible]	
FLOW RATE t/d	520	790	100	2281	23	305	2330	315	40	2531	2531	2226	671	4138	

C	GENERAL UPDATE	[illegible]	
B	RELEASED FOR FINAL REPORT	[illegible]	
A	RELEASED FOR DRAFT REPORT	[illegible]	
No	DESCRIPTION	DATE	BY

REVISIONS

REFERENCE DRAWINGS

PROCESS

SCALE NONE

ONTARIO ENERGY CORPORATION

TORONTO ONTARIO

KILBORN

TITLE

LIGNITE INDIRECT LIQUEFACTION

KOPPERS - TOTZEK

OPTION 2

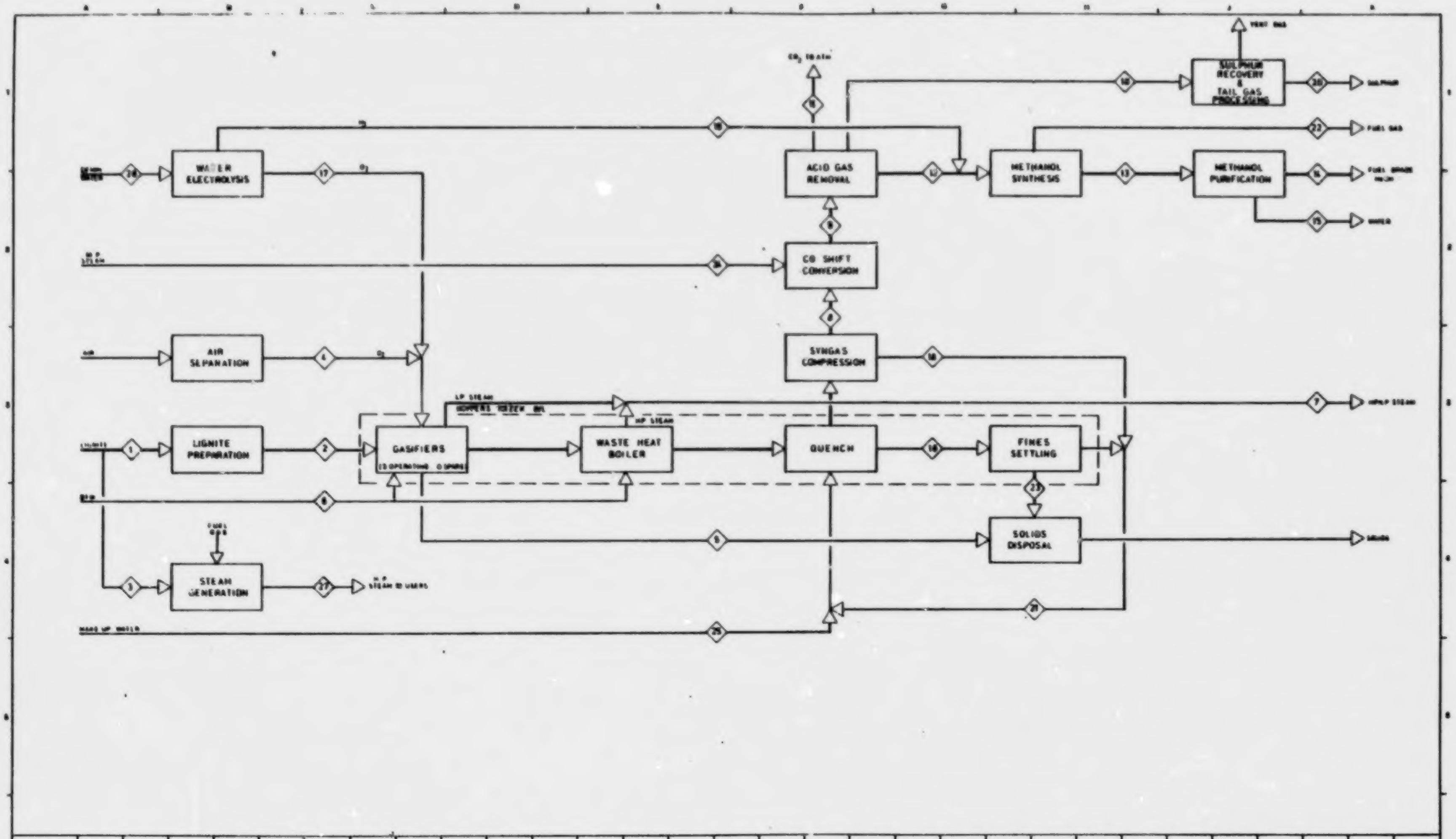
PRESSURE SWING ADSORPTION

3626

16

100 43-03

C


DEMIN WATER
26
WATER ELECTROLYSIS
17
O2
19
H.P. STEAM
24
AIR
AIR SEPARATION
4
O2
LIGNITE
1
LIGNITE PREPARATION
2
GASIFIERS
(3 OPERATING, 1 SPARE)
LP STEAM
KOPPERS TOTZEK BFL
HP STEAM
WASTE HEAT BOILER
BFW
6
FUEL GAS
STEAM GENERATION
3
27
H.P. STEAM TO USERS
MAKE UP WATER
25
5
QUENCH
SYNGAS COMPRESSION
8
CO SHIFT CONVERSION
9
ACID GAS REMOVAL
11
CO2 TO ATM
10
12
METHANOL SYNTHESIS
13
METHANOL PURIFICATION
14
FUEL GRADE MEOH
15
WATER
22
FUEL GAS
SULPHUR RECOVERY & TAIL GAS PROCESSING
VENT GAS
20
SULPHUR
16
FINES SETTLING
18
23
SOLIDS DISPOSAL
SOLIDS
21
7
HP/LP STEAM

STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
DESCRIPTION	AS RECEIVED LIGNITE TO PREP (35% MOIST)	LIGNITE TO GASIFIERS (10% MOIST)	LIGNITE TO STEAM PLANT (35% MOIST)	OXYGEN	SLAG (DRY BASIS)	BFW TO WASTE HEAT BOILER	HP STEAM (4000 KPa) & LP STEAM (350KPa)	RAW SYNGAS	SHIFTED SYNGAS TO AGR	ACID GAS TO CLAUS PLANT	CO_2 TO ATM	SYNGAS TO MEOH SYNTHESIS	RAW METHANOL	FUEL GRADE METHANOL	WATER	FINES TO SETTLING	OXYGEN	WATER	HYDROGEN	SULPHUR	RECYCLE WATER TO QUENCH	FUEL GAS	WET FINES TO DISPOSAL	STEAM TO CO SHIFT	MAKE-UP WATER	DEMIN WATER	HP SUPERH STEAM 400°C (10000 KPa)			
FLOW RATE t/d	4419	2643	2208	1206	235	3465	[illegible] 1260	3716	4578	38	2326	2216	2112	2080	102	780	758	150	95	35	437	187	470	1838	145	853	8475			

C	GENERAL UPDATE	23.2.84	
B	RELEASED FOR FINAL REPORT	10.8.83	
A	RELEASED FOR DRAFT REPORT	8.7.83	
No	DESCRIPTION		

REVISIONS

REFERENCE DRAWINGS

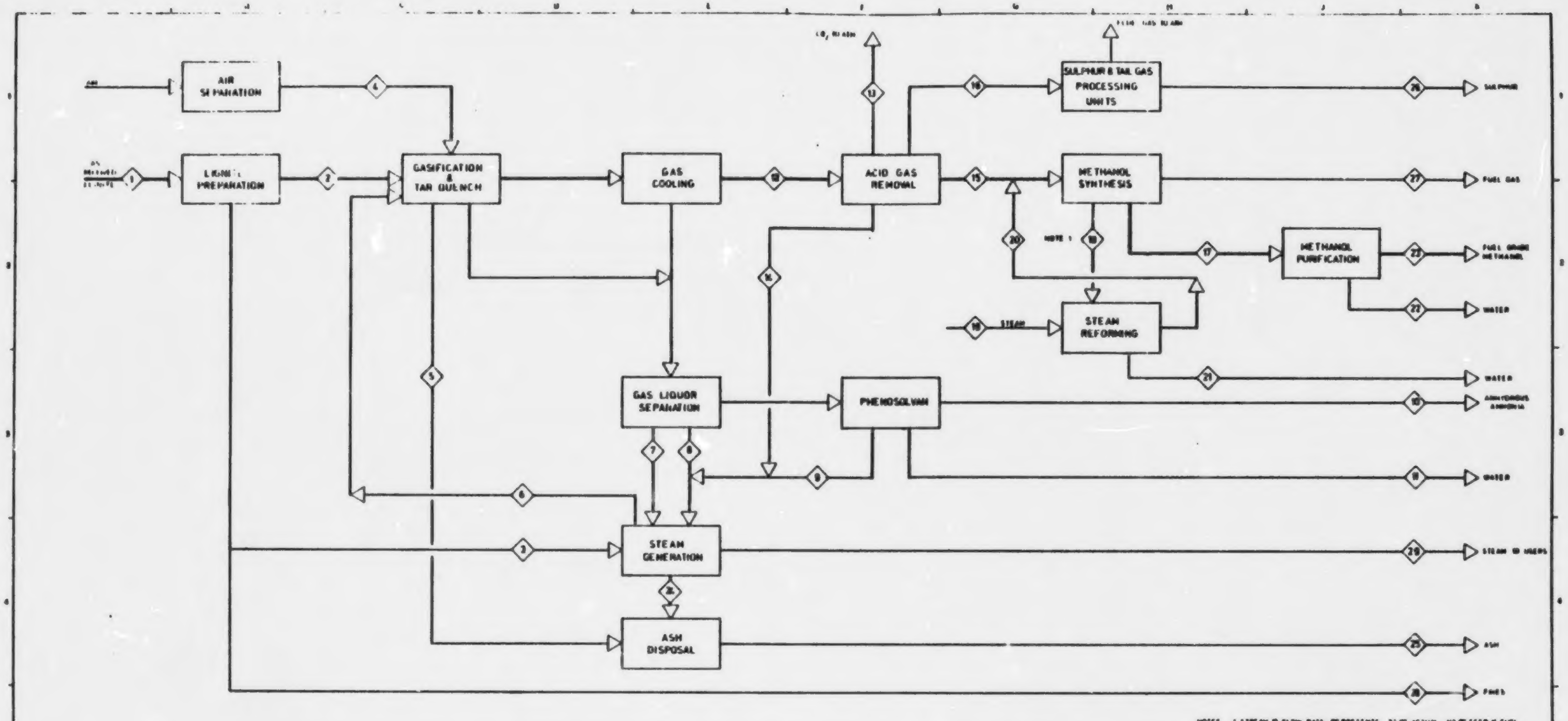
PROCESS
SCALE NONE
AS
AS
JWC
TRM

CLIENT ONTARIO ENERGY CORPORATION
LOCATION TORONTO ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
KOPPERS - TOTZEK
OPTION 3
ELECTROLYSIS

3424 16
DRAWING NUMBER
100-43-04 C


AIR
AIR SEPARATION
LIGNITE PREPARATION
GASIFICATION & TAR QUENCH
GAS COOLING
ACID GAS REMOVAL
CO2 TO ATM
SULPHUR & TAIL GAS PROCESSING UNITS
FLUE GAS TO ATM
METHANOL SYNTHESIS
NOTE 1
METHANOL PURIFICATION
STEAM
STEAM REFORMING
GAS LIQUOR SEPARATION
PHENOSOLVAN
STEAM GENERATION
ASH DISPOSAL
SULPHUR
FUEL GAS
FUEL GRADE METHANOL
WATER
WATER
ANHYDROUS AMMONIA
WATER
STEAM TO USERS
ASH
FINES
1
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NOTES 1 STREAM 18 FLOW RATE REPRESENTS 2/3 OF ACTUAL. 1/3 OF FEED IS FUEL

STREAM No	DESCRIPTION	FLOW RATE t/d
1	RUN OF MINE LIGNITE (35% MOIST.)	8459
2	LIGNITE TO GASIFIER (30% MOIST.)	4985
3	LIGNITE FINES TO STEAM PLANT (35% MOIST.)	493
4	OXYGEN	994
5	SLAG TO ASH DISPOSAL	657
6	STEAM TO GASIFIER	3811
7	TARS TO FUEL	138
8	TAR OILS TO FUEL	34
9	PHENOLS TO FUEL	23
10	ANHYD AMMONIA	35
11	WATER	3310
12	RAW SYNGAS TO AGR	5613
13	CO_2 TO ATM	3580
14	NAPHTHA TO FUEL	26
15	SYNGAS TO METHANOL SYNTHESIS	1958
16	ACID GAS TO CLAUS PLANT	52
17	RAW MeOH TO PURIF	2142
18	PURGE GAS TO REFORMER	339
19	STEAM TO REFORMER	1012
20	SYNGAS FROM REFORMER	826
21	WATER	525
22	WATER	133
23	FUEL GRADE MeOH	2010
24	ASH FROM STEAM PLANT	52
25	ASH TO LANDFILL	709
26	SULPHUR	49
27	FUEL GAS	303
28	LIGNITE FINES (35% MOIST.)	1822
29	HP STEAM 10.0 MPa	4200
30		

REFERENCE DRAWINGS

REVISIONS

No.	DESCRIPTION	DATE
C	GENERAL UPDATE	10/84
B	RELEASED FOR FINAL REPORT	08 81
A	RELEASED FOR DRAFT REPORT	07 81

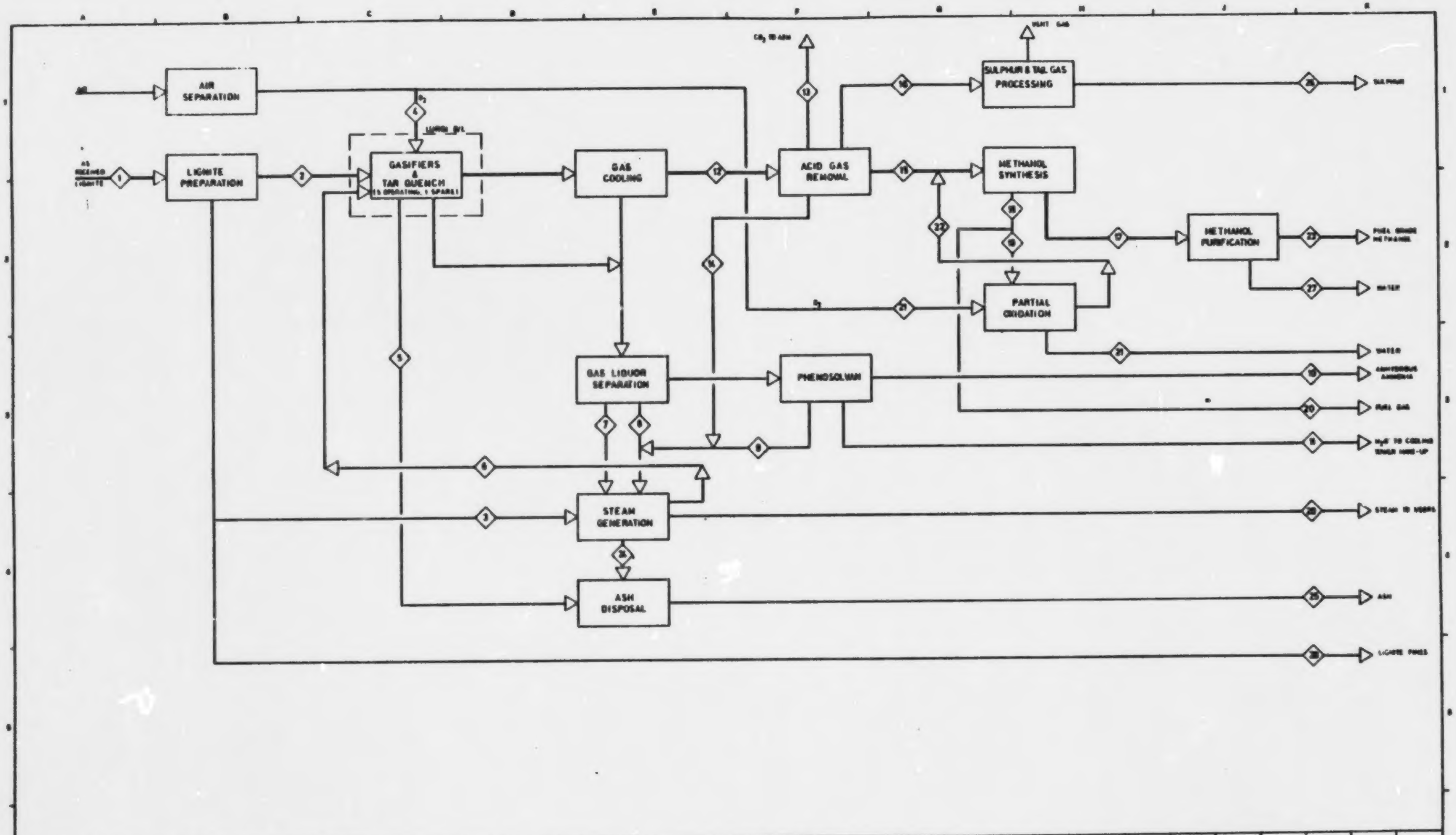
DESIGN PROCESS
SCALE NONE
DESIGNED AS
DRAWN AS
CHECKED JWC
APPROVED [illegible]

CLIENT ONTARIO ENERGY CORPORATION
LOCATION TORONTO ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
LURGI
BASE CASE

PROJECT No 3426
DRAWING NUMBER 200 43 01
REV C


AIR
AIR SEPARATION
AS RECEIVED LIGNITE
LIGNITE PREPARATION
O_2
LURGI B/L
GASIFIERS & TAR QUENCH (5 OPERATING, 1 SPARE)
GAS COOLING
GAS LIQUOR SEPARATION
STEAM GENERATION
ASH DISPOSAL
ACID GAS REMOVAL
CO_2 TO ATM
PHENOSOLVAN
SULPHUR & TAIL GAS PROCESSING
VENT GAS
METHANOL SYNTHESIS
PARTIAL OXIDATION
METHANOL PURIFICATION
SULPHUR
FUEL GRADE METHANOL
WATER
WATER
ANHYDROUS AMMONIA
FUEL GAS
H_2O TO COOLING TOWER MAKE-UP
STEAM TO USERS
ASH
LIGNITE FINES
1
2
3
4
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12
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14
16
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17
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19
20
21
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11

STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
DESCRIPTION	RUN OF MINE LIGNITE (35% MOIST)	LIGNITE TO GASIFIERS (30% MOIST)	LIGNITE FINES TO STEAM PLANT (35% MOIST)	OXYGEN	SLAG TO ASH DISPOSAL	STEAM TO GASIFIERS	TARS TO FUEL	TAR OILS TO FUEL	PHENOLS TO FUEL	ANHYD AMMONIA	WATER	RAW SYNGAS TO AGR	CO_2 TO ATM	NAPHTHA TO FUEL	SYNGAS TO METHANOL SYNTHESIS
FLOW RATE	8430	4805	860	984	656	3830	130	34	23	35	3310	564	3580	24	1858

STREAM No	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
DESCRIPTION	ACID GAS TO CLAUS PLANT	RAW MeOH TO PURIF	PURGE GAS TO REFORMER	POX FEED	FUEL GAS	OXYGEN	POX EFFLUENT GAS	FUEL GRADE MeOH	ASH FROM STEAM PLANT	ASH TO LANDFILL	SULPHUR	WATER	EXCESS FINES	STEAM TO USERS	
FLOW RATE	52	2005	607	388	238	388	736	2010	178	834	48	76	1165	4086	

No	DESCRIPTION	DATE	BY
C	GENERAL UPDATE	23.2.84	
B	RELEASED FOR FINAL REPORT	10.8.83	
A	RELEASED FOR DRAFT REPORT	8.7.83	

REVISIONS

REFERENCE DRAWINGS

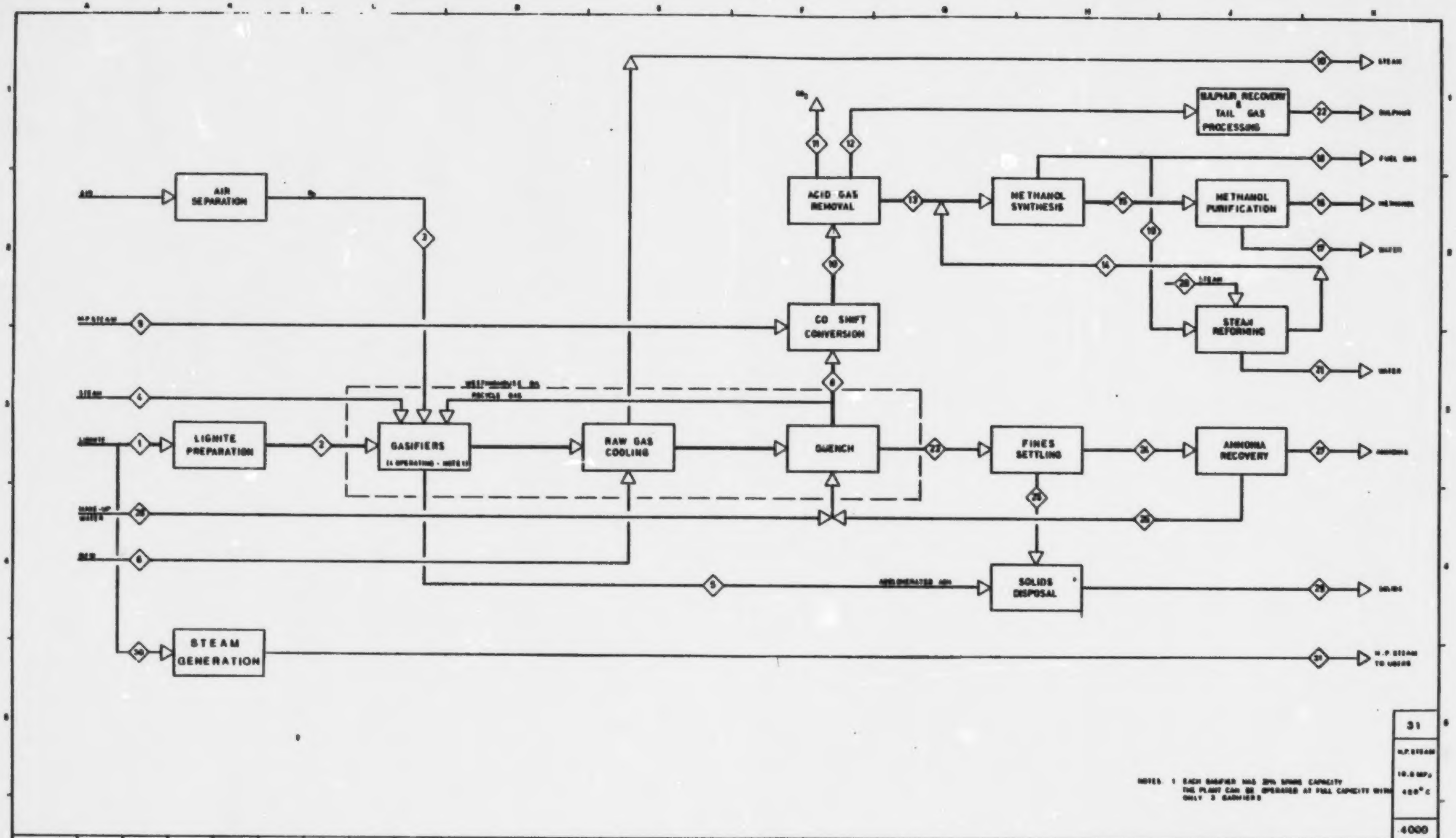
DESIGN PROCESS
SCALE NONE
DRAWN AS
CHECKED AS
APPROVED JWC

CLIENT ONTARIO ENERGY CORPORATION
LOCATION TORONTO, ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
LURGI
OPTION "A"
PARTIAL OXIDATION

JOB No 3424 16
DRAWING NUMBER 200-43-02
REV C


AIR
AIR SEPARATION
O_2
M.P. STEAM
STEAM
LIGNITE
MAKE-UP WATER
BFW
LIGNITE PREPARATION
GASIFIERS
(4 OPERATING - NOTE 1)
WESTINGHOUSE OIL RECYCLE GAS
RAW GAS COOLING
QUENCH
CO SHIFT CONVERSION
ACID GAS REMOVAL
CO_2
METHANOL SYNTHESIS
METHANOL PURIFICATION
STEAM REFORMING
SULPHUR RECOVERY & TAIL GAS PROCESSING
FINES SETTLING
AMMONIA RECOVERY
AGGLOMERATED ASH
SOLIDS DISPOSAL
STEAM GENERATION
STEAM
SULPHUR
FUEL GAS
METHANOL
WATER
WATER
AMMONIA
SOLIDS
H.P. STEAM TO USERS

NOTES: 1 EACH GASIFIER HAS 25% SPARE CAPACITY
THE PLANT CAN BE OPERATED AT FULL CAPACITY WITH ONLY 3 GASIFIERS

STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP (35% MOISTURE)	LIGNITE TO GASIFIERS (8% MOIST.)	OXYGEN TO GASIFIERS	STEAM TO GASIFIERS (2100 kPa)	AGGLOM ASH (DRY BASIS)	B F W	HP STEAM 4.3 MPa 430°C	RAW SYNGAS TO SHIFT CONVERTOR	M P STEAM	SHIFTED SYNTHESIS GAS	CO_2 TO ATM	ACID GAS TO CLAUS PLANT	TREATED SYNTHESIS GAS	SYNGAS FROM REFORMING	RAW METHANOL TO PURIF
FLOW RATE t/h	5872	3980	1671	668	655	3883	3698	5492	1593	5892	3720	51	2221	254	2212

STREAM No	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
DESCRIPTION	FUEL GRADE METHANOL	WATER	FUEL GAS	REFORMER FEED	STEAM TO REFORMER	WATER	SULPHUR	UNTREATED WATER	QUENCH WATER CONTAINING FINES	SLURRY TO SOLIDS DISPOSAL	TREATED WATER RECYCLE	ANHYDROUS AMMONIA	MAKE-UP WATER	SOLIDS TO DISPOSAL	LIGNITE TO STEAM GENERATION	H.P. STEAM 10.0 MPa 480°C
FLOW RATE t/h	2010	202	204	104	311	161	48	402	255	47	348	7	62	782	1284	4000

REFERENCE DRAWINGS

NO.	DESCRIPTION	DATE	BY
C	GENERAL UPDATE	21 2 84	
B	RELEASED FOR FINAL REPORT	10 0 83	
A	RELEASED FOR DRAFT REPORT	0 7 83	

REVISIONS

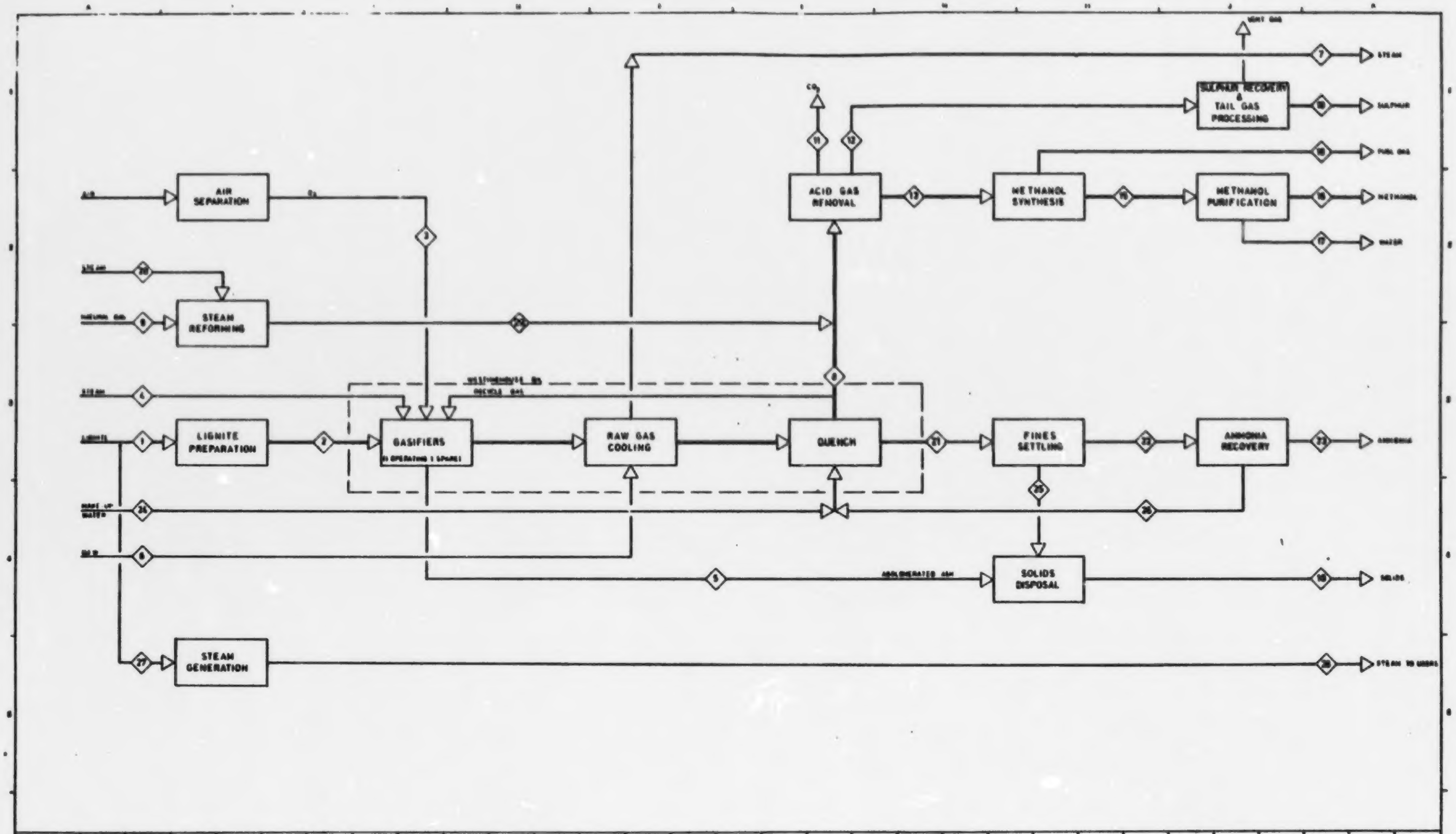
SECTION PROCESS
SCALE NONE
DESIGNED AS
DRAWN AS
CHECKED JWC
APPROVED JHM

CLIENT ONTARIO ENERGY CORPORATION
LOCATION TORONTO ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
WESTINGHOUSE
BASE CASE

JOB NO.
3424 15
DRAWING NUMBER
300 43-01
REV C


AIR
AIR SEPARATION
O_2
STEAM
NATURAL GAS
STEAM REFORMING
STEAM
LIGNITE
LIGNITE PREPARATION
MAKE UP WATER
BFW
STEAM GENERATION
GASIFIERS
(3 OPERATING 1 SPARE)
WESTINGHOUSE BS RECYCLE GAS
RAW GAS COOLING
QUENCH
AGGLOMERATED ASH
ACID GAS REMOVAL
CO_2
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
VENT GAS
FINES SETTLING
AMMONIA RECOVERY
SOLIDS DISPOSAL
STEAM
SULPHUR
FUEL GAS
METHANOL
WATER
AMMONIA
SOLIDS
STEAM TO USERS
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STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP (35% MOISTURE)	LIGNITE TO GASIFIER (8% MOISTURE)	OXYGEN TO GASIFIER	STEAM TO GASIFIER (2100 kPa)	AGGLOM ASH (DRY BASIS)	B F W	HP STEAM 6.2 MPa 435°C	RAW SYNGAS TO SHIFT CONVERSION	NATURAL GAS	SOLIDS TO DISPOSAL	CO_2 TO ATM	ACID GAS TO CLAUS PLANT	TREATED SYNTHESIS GAS	SYNGAS FROM REFORMING	RAW METHANOL TO PURIF
FLOW RATE t/d	1534	993	437	176	171	971	926	1434	737	181	508	13	2858	1743	2488

STREAM No	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30
DESCRIPTION	FUEL GRADE METHANOL	WATER	FUEL GAS	SULPHUR	STEAM TO REFORMER	UNTREATED WATER	QUENCH WATER CONTAINING FINES	ANHYDROUS AMMONIA	MAKE-UP WATER	SLURRY TO SOLIDS DISPOSAL	TREATED WATER RECYCLE	LIGNITE TO STEAM PLANT	HP STEAM	REFORMED GAS	
FLOW RATE t/d	2227	272	156	12	1805	105	95	2	18	10	83	530	1687	1742	

REFERENCE DRAWINGS

No	DESCRIPTION	DATE	BY
C	GENERAL UPDATE	83.2.04	
B	RELEASED FOR FINAL REPORT	83 8 03	
A	RELEASED FOR DRAFT REPORT	8 7 83	

REVISIONS

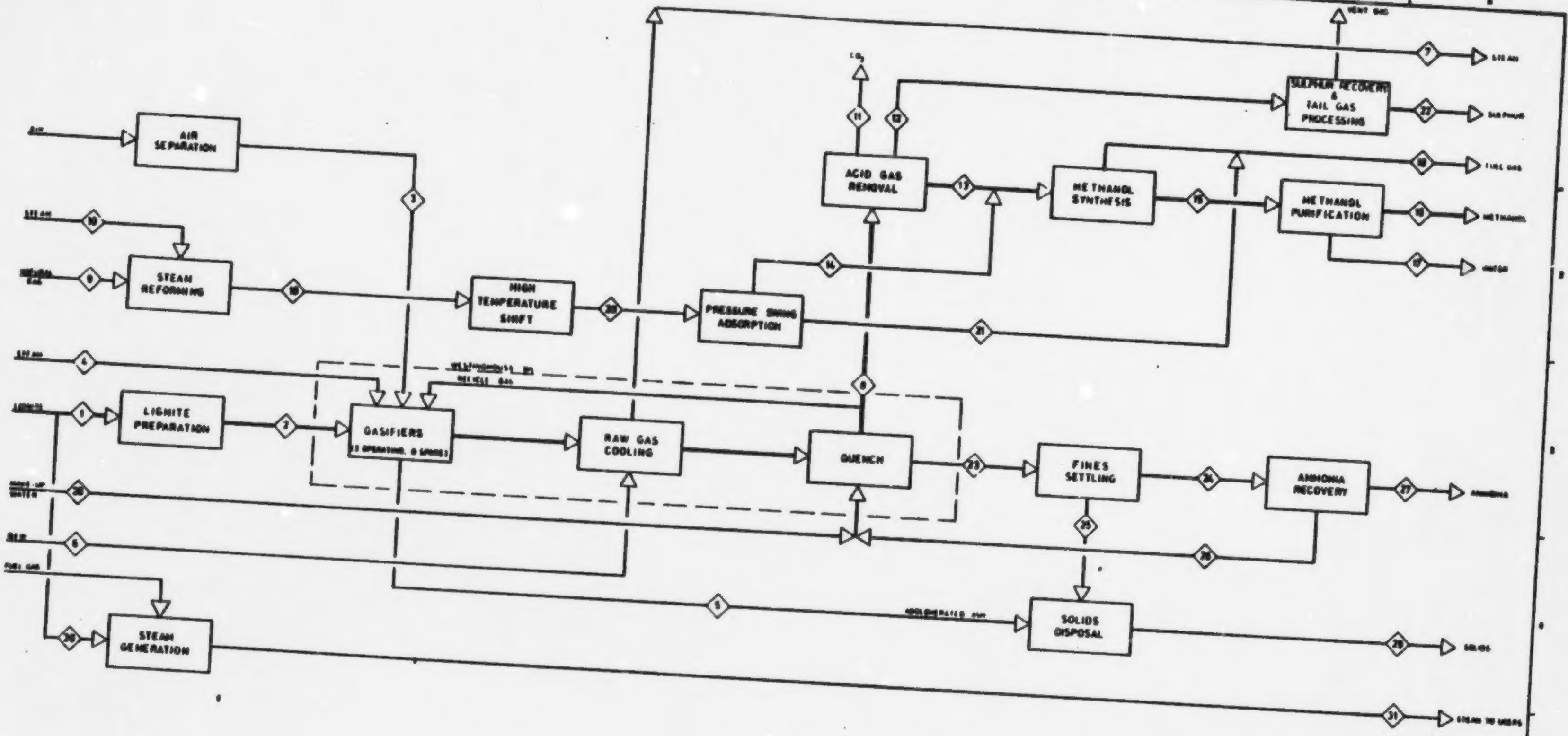
SECTION PROCESS
SCALE NONE
DESIGNED AS
DRAWN AS
CHECKED JWC
APPROVED

CLIENT ONTARIO ENERGY CORPORATION
LOCATION TORONTO ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
WESTINGHOUSE
OPTION I
STEAM-METHANE REFORMING

PROJECT NO 3434 | SEQUENCE 15
DRAWING NUMBER 300-43-02 REV C


AIR
AIR SEPARATION
STEAM
NATURAL GAS
STEAM REFORMING
HIGH TEMPERATURE SHIFT
PRESSURE SWING ADSORPTION
STEAM
LIGNITE
LIGNITE PREPARATION
GASIFIERS (13 OPERATING, 0 SPARES)
WESTINGHOUSE OR RECYCLE GAS
RAW GAS COOLING
QUENCH
MAKE-UP WATER
BLR
FUEL GAS
STEAM GENERATION
AGGLOMERATED ASH
SOLIDS DISPOSAL
FINES SETTLING
AMMONIA RECOVERY
ACID GAS REMOVAL
CO2
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
VENT GAS
STEAM
SULPHUR
FUEL GAS
METHANOL
WATER
AMMONIA
SOLIDS
STEAM TO USERS
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STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP (35% MOISTURE)	LIGNITE TO GASIFIERS (5% MOISTURE)	OXYGEN TO GASIFIERS	STEAM TO GASIFIERS (3100 kPa)	AGGLOM ASH (DRY BASIS)	B.F.W.	HP STEAM 4.2 MPa 425°C	RAW GAS	NATURAL GAS	STEAM	CO_2 TO ATM	ACID GAS TO CLAUS PLANT	TREATED SYNTHESIS GAS	HYDROGEN	RAW METHANOL TO PURIF
FLOW RATE (t/d)	3068	1985	873	348	342	1042	1850	2869	641	1412	278	27	2564	249	2448

STREAM No	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
DESCRIPTION	FUEL GRADE METHANOL	WATER	FUEL GAS	REFORMER EFFLUENT	SHIFT EFFLUENT	REGEN GAS	SULPHUR	UNTREATED WATER	QUENCH WATER CONTAINING FINES	SLURRY TO SOLIDS DISPOSAL	TREATED WATER RECYCLE	ANHYDROUS AMMONIA	MAKE-UP WATER	SOLIDS TO DISPOSAL	LIGNITE TO STEAM PLANT	HP STEAM
FLOW RATE (t/d)	2144	305	2048	2052	2052	1803	25	210	181	19	187	4	32	381	678	2836

REFERENCE DRAWINGS

	DESCRIPTION	DATE
C	GENERAL UPDATE	16.2.84
B	RELEASED FOR FINAL REPORT	10.6.83
A	RELEASED FOR DRAFT REPORT	6.7.83

REVISIONS

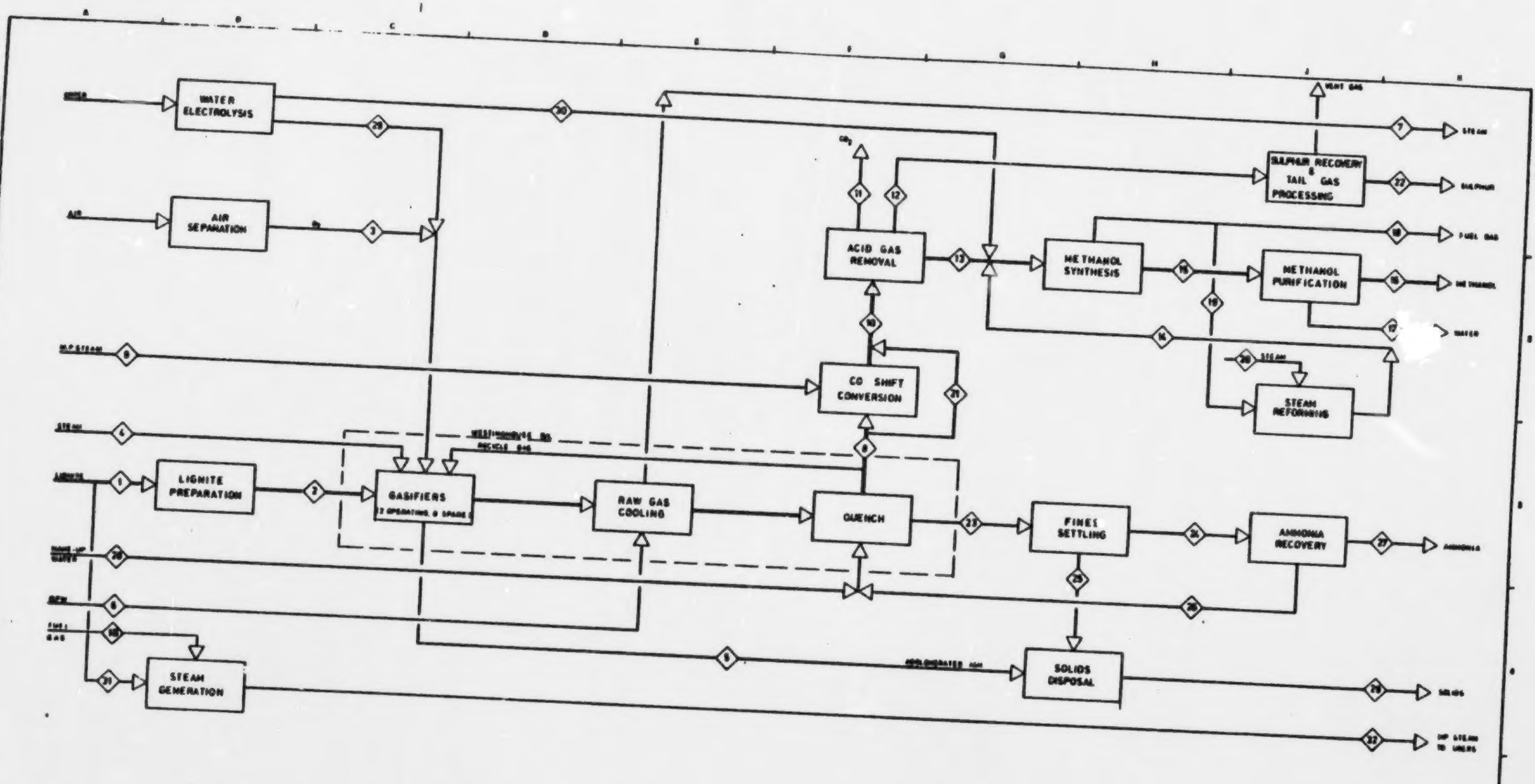
SECTION PROCESS
SCALE NONE
DESIGNED AS
DRAWN AS
CHECKED JMC
APPROVED JMH

CLIENT ONTARIO ENERGY CORPORATION
LOCATION TORONTO ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
WESTINGHOUSE
OPTION 2
STEAM METHANE REF./PRESSURE SWING ADSORPTION

PROJECT No 3426
DRAWING No 300 43-03
REV C


WATER
WATER ELECTROLYSIS
AIR
AIR SEPARATION
M.P. STEAM
STEAM
LIGNITE
LIGNITE PREPARATION
GASIFIERS (2 OPERATING, 0 SPARE)
WESTINGHOUSE RG RECYCLE GAS
RAW GAS COOLING
QUENCH
CO SHIFT CONVERSION
ACID GAS REMOVAL
CO_2
METHANOL SYNTHESIS
METHANOL PURIFICATION
STEAM REFORMING
SULPHUR RECOVERY & TAIL GAS PROCESSING
VENT GAS
FINES SETTLING
AMMONIA RECOVERY
SOLIDS DISPOSAL
STEAM GENERATION
MAKE-UP WATER
BFW
FUEL GAS
AGGLOMERATED ASH
STEAM
SULPHUR
FUEL GAS
METHANOL
WATER
AMMONIA
SOLIDS
HP STEAM TO USERS
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STREAM No	1	2	3	4	5	6	7	8	9	10	11
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP (35% MOISTURE)	LIGNITE TO GASIFIERS (MOISTURE)	OXYGEN FROM AIR SEPARATION	STEAM TO GASIFIERS (3100 kPa)	AGGLOM ASH (DRY BASIS)	B F W	HP STEAM 4.2 MPa 425°C	RAW GAS	M.P. STEAM	SHIFTED SYNTHESIS GAS	CO_2 TO ATM
FLOW RATE[illegible]	4448	2878	396	500	497	2933	2784	4868	778	4390	2029

STREAM No	12	13	14	15	16	17	18	19	20	21	22
DESCRIPTION	ACID GAS TO CLAUS PLANT	TREATED SYNTHESIS GAS	SYNGAS FROM REFORMING	RAW METHANOL TO PURIF	FUEL GRADE METHANOL	WATER	FUEL GAS	REFORMER FEED	STEAM TO REFORMER	CO SHIFT BY-PASS	SULPHUR
FLOW RATE[illegible]	30	2231	192	2250	2090	248	246	70	226	1037	38

STREAM No	23	24	25	26	27	28	29	30	31	32
DESCRIPTION	UNTREATED WATER	QUENCH WATER CONTAINING FINES	SLURRY TO SOLIDS DISPOSAL	TREATED WATER RECYCLE	ANHYDROUS AMMONIA	MAKE-UP WATER	ELECTROLYTIC OXYGEN	ELECTROLYTIC HYDROGEN	LIGNITE TO STEAM PLANT	HP STEAM 10.0 MPa 480°C
FLOW RATE[illegible]	383	276	29	271	3	37	600	101	381	2801

C	GENERAL UPDATE	10.2.81
B	RELEASED FOR FINAL REPORT	10.0.83
A	RELEASED FOR DRAFT REPORT	5.7.80
No	DESCRIPTION	DATE

REVISIONS

REFERENCE DRAWINGS

PROCESS

SCALE NONE

DRAWN AS

CHECKED AS

ONTARIO ENERGY CORPORATION

TORONTO ONTARIO

KILBORN

TITLE

LIGNITE INDIRECT LIQUEFACTION

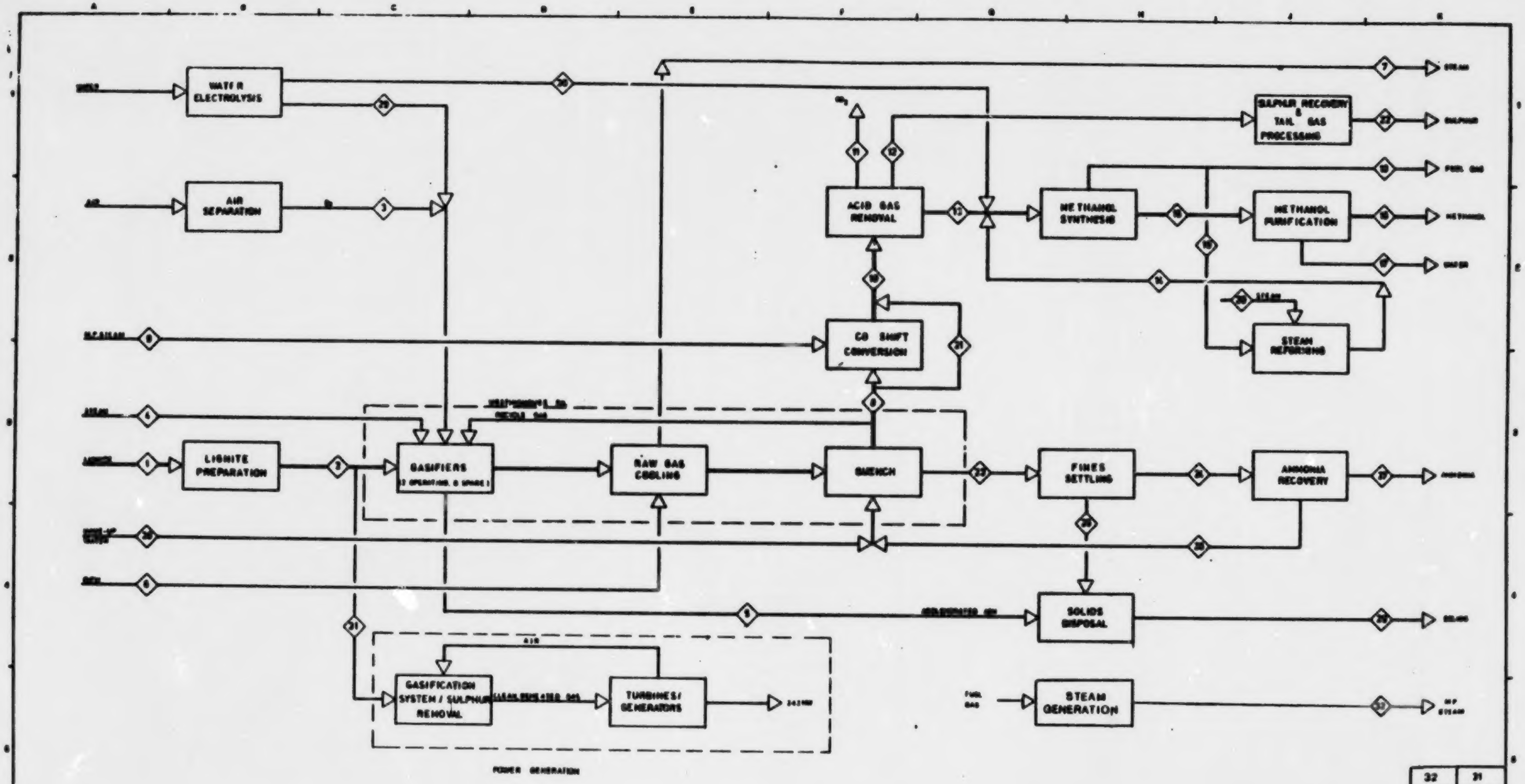
WESTINGHOUSE

OPTION 3 - ELECTROLYSIS

3426 16

DRAWING NUMBER

300-43-04 C


WATER
AIR
MP STEAM
STEAM
LIGNITE
MAKE-UP WATER
BFW
WATER ELECTROLYSIS
AIR SEPARATION
LIGNITE PREPARATION
GASIFIERS
RAW GAS COOLING
QUENCH
CO SHIFT CONVERSION
ACID GAS REMOVAL
CO_2
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
STEAM REFORMING
STEAM
FINES SETTLING
AMMONIA RECOVERY
SOLIDS DISPOSAL
FUEL GAS
STEAM GENERATION
GASIFICATION SYSTEM / SULPHUR REMOVAL
AIR
TURBINES/ GENERATORS
POWER GENERATION
STEAM
SULPHUR
FUEL GAS
METHANOL
WATER
AMMONIA
SOLIDS
MP STEAM

STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP	LIGNITE TO GASIFIERS	OXYGEN FROM AIR SEPARATION	STEAM TO GASIFIERS	ASH ... (DRY BASIS)	B F W	HP STEAM	RAW SYNGAS TO SHIFT CONVERSION	MP STEAM	SHIFTED SYNTHESIS GAS	CO_2 TO ATM	ACID GAS TO CLAUS PLANT	TREATED SYNTHESIS GAS	SYNGAS FROM REFORMING	RAW METHANOL TO PURIF	FUEL GRADE METHANOL
FLOW RATE	8680	5568	396	505	497	2833	2794	4148	778	4390	2023	38	2331	192	2250	2010

STREAM No	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32
DESCRIPTION	WATER	FUEL GAS	REFORMER FEED	STEAM TO REFORMER	CO SHIFT BY PASS	SULPHUR	UNTREATED WATER	QUENCH WATER CONTAINING FINES	SLURRY TO SOLIDS DISPOSAL	TREATED WATER RECYCLE	ANHYDROUS AMMONIA	MAKE-UP WATER	ELECTROLYTIC OXYGEN	ELECTROLYTIC HYDROGEN	LIGNITE TO POWER GENER	MP STEAM
FLOW RATE	240	348	70	235	1037	35	303	274	29	271	3	37	801	101	2880	900

C	GENERAL UPDATE	21 2 84
B	RELEASED FOR FINAL REPORT	30 8 83
A	RELEASED FOR DRAFT REPORT	8 7 83

REVISIONS

REFERENCE DRAWINGS

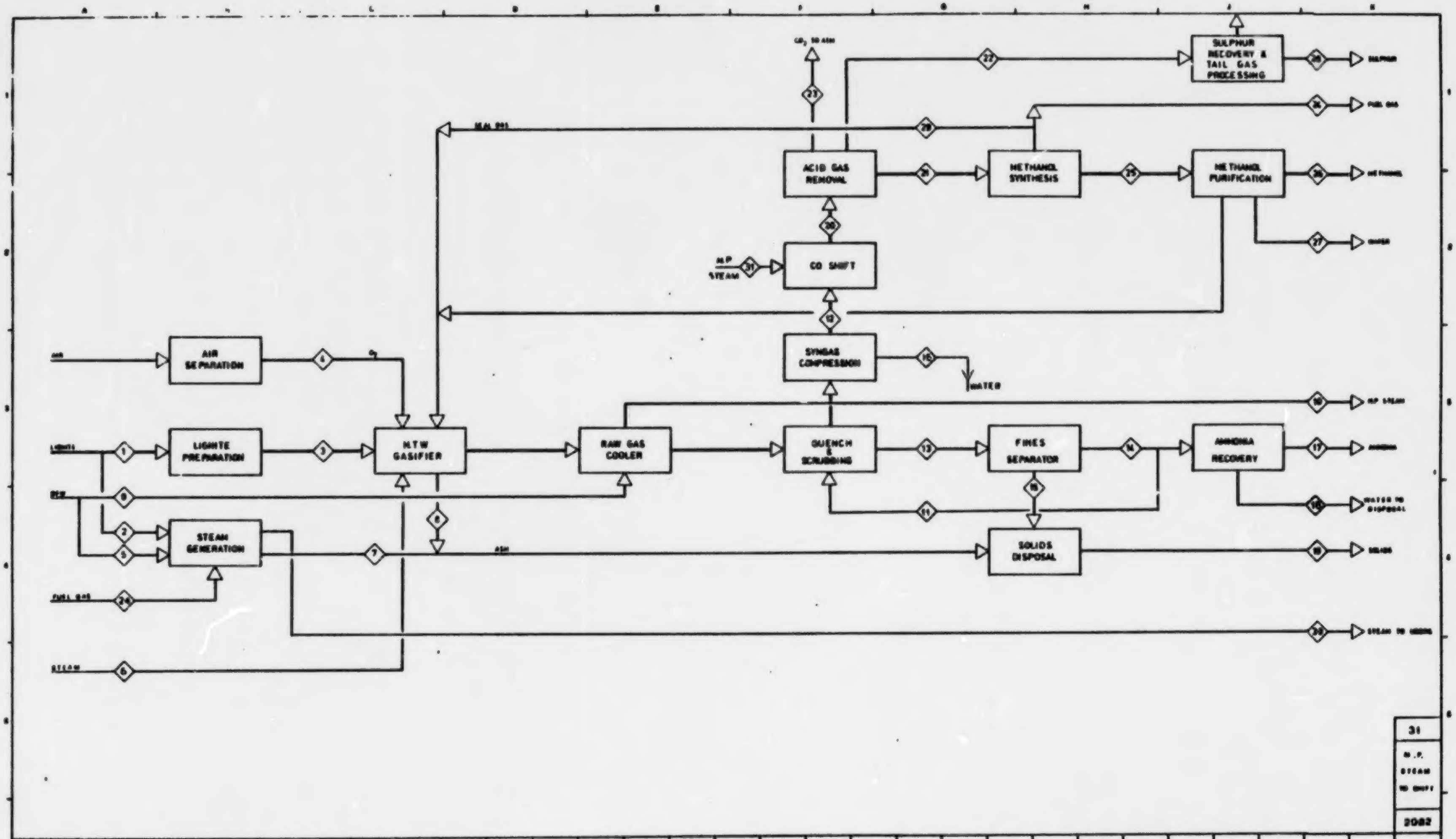
PROCESS
SCALE NONE

ONTARIO ENERGY CORPORATION
LOCATION TORONTO, ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
WESTINGHOUSE
OPTION 4
CO-GENERATION

3436 15
DRAWING NUMBER
300-42-05 C


AIR
AIR SEPARATION
O_2
LIGNITE
LIGNITE PREPARATION
BFW
STEAM GENERATION
FUEL GAS
STEAM
HTW GASIFIER
SEAL GAS
ASH
RAW GAS COOLER
QUENCH & SCRUBBING
SYNGAS COMPRESSION
WATER
CO SHIFT
HP STEAM
ACID GAS REMOVAL
CO_2 TO ATM
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
FINES SEPARATOR
AMMONIA RECOVERY
SOLIDS DISPOSAL
SULPHUR
FUEL GAS
METHANOL
WATER
HP STEAM
AMMONIA
WATER TO DISPOSAL
SOLIDS
STEAM TO USERS
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STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP [illegible] MOIST.	AS RECEIVED LIGNITE TO STEAM PLANT [illegible] MOIST.	LIGNITE TO GASIFIER [illegible] MOIST.	OXYGEN TO GASIFIER	BFW TO STEAM PLANT	STEAM TO GASIFIER	ASH FROM STEAM PLANT TO DISPOSAL	ASH FROM GASIFIER TO DISPOSAL	BFW	HP STEAM	QUENCH WATER	RAW SYNGAS TO CO SHIFT	QUENCH WATER WITH FINES	SEPARATED WATER TO NH_3 RECOVERY	SLUDGE TO SOLIDS DISPOSAL
FLOW RATE	5606	1043	2504	2200	7140	960	200	551	3104	2967	700	5318	1760	1032	128

STREAM No	16	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31
DESCRIPTION	WATER FROM COMPRESSION	ANHYDROUS AMMONIA	WATER TO DISPOSAL	SOLIDS TO LANDFILL	SHIFTED SYNGAS TO ACID GAS REMOVAL	SYNGAS TO METHANOL SYNTHESIS	ACID GAS TO CLAUS PLANT	CO_2 TO ATM	FUEL GAS	RAW METHANOL TO PURIFICATION	FUEL GRADE METHANOL	WATER	SULPHUR	SEAL GAS RECYCLE TO GASIFIER	HP STEAM	H.P. STEAM TO SHIFT
FLOW RATE	19	12	920	870	6360	2386	60	3904	181	2177	2080	182	56	181	7620	2082

	DESCRIPTION	DATE
C	GENERAL UPDATE	23.8.84
B	RELEASED FOR FINAL REPORT	19.8.83
A	RELEASED FOR DRAFT REPORT	8.7.83

REVISIONS

REFERENCE DRAWINGS

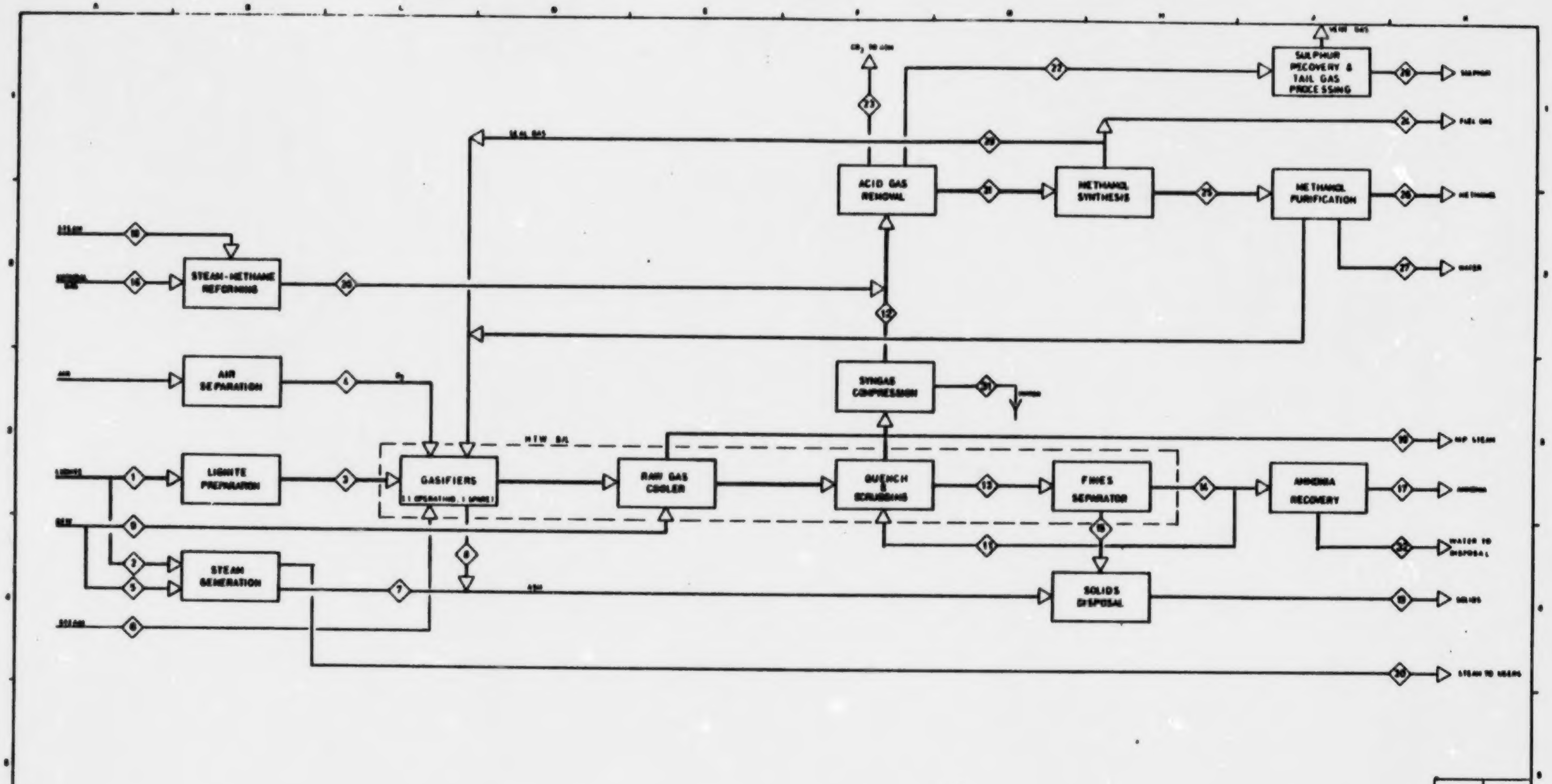
PROCESS
SCALE NONE
DESIGNED AS
DRAWN AS
CHECKED JWC
APPROVED JDM

ONTARIO ENERGY CORPORATION
TORONTO, ONTARIO

KILBORN

TITLE
LIGHITE INDIRECT LIQUEFACTION
HIGH TEMPERATURE WINKLER
BASE CASE

3424 | 10
DRAWING NUMBER
400-43-01 | C


STEAM-METHANE REFORMING
AIR SEPARATION
LIGNITE PREPARATION
STEAM GENERATION
GASIFIERS (1 OPERATING, 1 SPARE)
HTW B/L
RAW GAS COOLER
QUENCH & SCRUBBING
FINES SEPARATOR
SOLIDS DISPOSAL
SYNGAS COMPRESSION
ACID GAS REMOVAL
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
AMMONIA RECOVERY
STEAM
NATURAL GAS
AIR
LIGNITE
BFW
SEAL GAS
ASH
O_2
CO_2 TO ATM
VENT GAS
SULPHUR
FUEL GAS
METHANOL
WATER
HP STEAM
AMMONIA
WATER TO DISPOSAL
SOLIDS
STEAM TO USERS

STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP ([illegible]% MOIST)	AS RECEIVED LIGNITE TO STEAM PLANT ([illegible]% MOIST)	LIGNITE TO GASIFIERS ([illegible]% MOIST)	OXYGEN TO GASIFIERS	BFW TO STEAM PLANT	STEAM TO GASIFIERS	ASH FROM STEAM PLANT TO DISPOSAL	ASH FROM GASIFIER TO DISPOSAL	BFW	HP STEAM	QUENCH WATER	RAW SYNGAS	QUENCH WATER WITH FINES	SEPARATED WATER TO NH_3 RECOVERY	SLUDGE TO SOLIDS DISPOSAL	NATURAL GAS
FLOWRATE	1917	766	1188	732	1593	336	79	188	1039	890	180	1818	200	424	61	877

STREAM No	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32
DESCRIPTION	ANHYDROUS AMMONIA	STEAM	SOLIDS TO LANDFILL	REFORMED GAS	SYNGAS TO METHANOL SYNTHESIS	ACID GAS TO CLAUS PLANT	CO_2 TO ATM	FUEL GAS	RAW METHANOL TO PURIFICATION	FUEL GRADE METHANOL	WATER	SULPHUR	SEAL GAS RECYCLE TO GASIFIER	HP STEAM TO USERS	WATER	WATER TO DISPOSAL
FLOWRATE	4	1193	329	3070	3024	20	843	75	2951	2780	[illegible]7	18	14	2446	7	240

	DESCRIPTION		
C	GENERAL UPDATE	23 2 84	
B	RELEASED FOR FINAL REPORT	10 8 83	
A	RELEASED FOR DRAFT REPORT	6 7 83	
No	DESCRIPTION	DATE	BY

REVISIONS

REFERENCE DRAWINGS

PROCESS

SCALE NONE

AS

AS

CHECKED JWC

CLIENT ONTARIO ENERGY CORPORATION

LOCATION TORONTO ONTARIO

KILBORN

TITLE

LIGNITE INDIRECT LIQUEFACTION

HIGH TEMPERATURE WINKLER

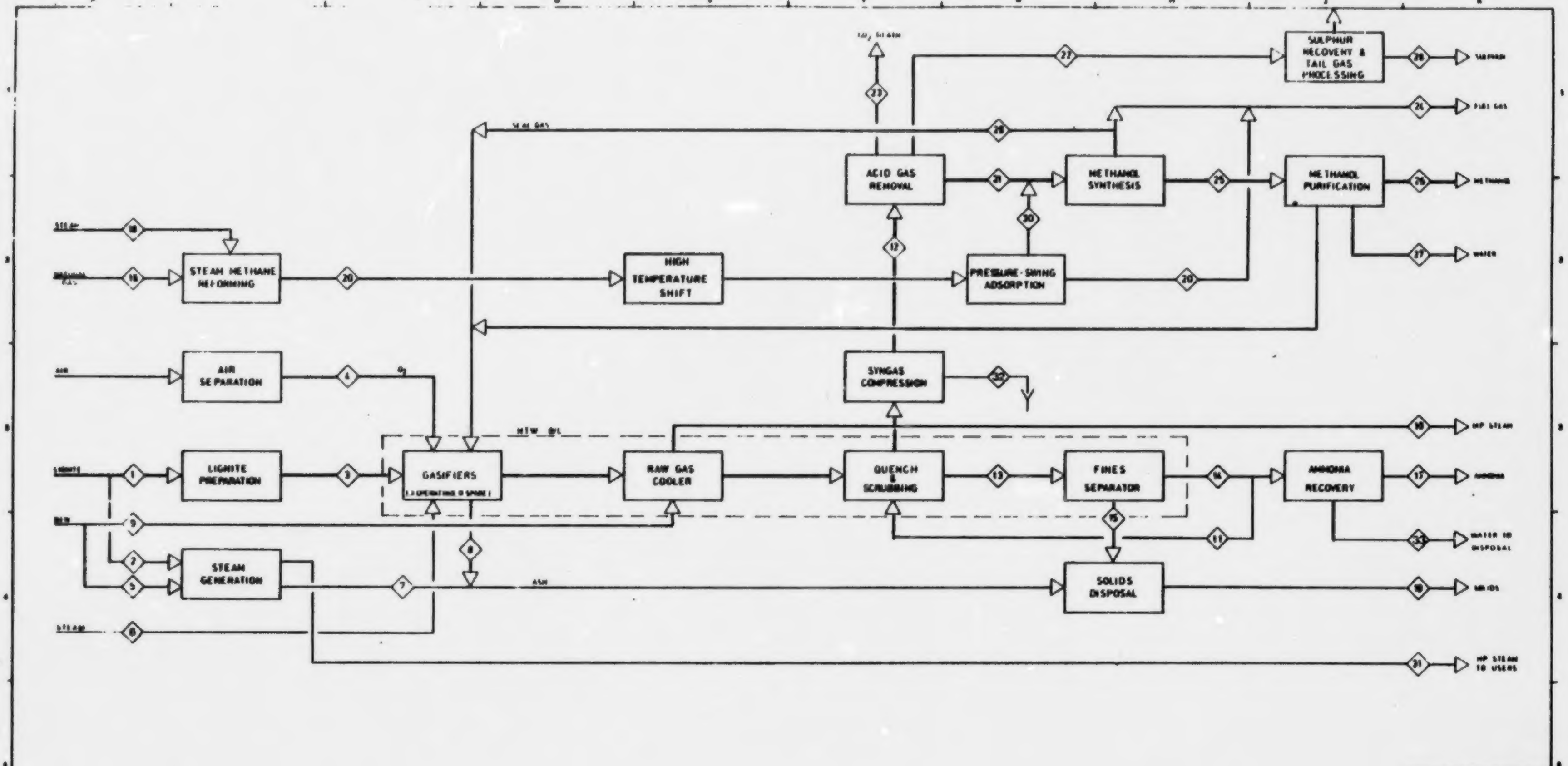
OPTION 1

STEAM-METHANE REFORMING

PROJECT No 3426

DRAWING NUMBER 400-43-02

REV C


STEAM
NATURAL GAS
AIR
LIGNITE
BFW
STEAM
STEAM METHANE REFORMING
AIR SEPARATION
LIGNITE PREPARATION
STEAM GENERATION
GASIFIERS (3 OPERATING 0 SPARE)
HTW
RAW GAS COOLER
QUENCH & SCRUBBING
FINES SEPARATOR
SOLIDS DISPOSAL
SYNGAS COMPRESSION
HIGH TEMPERATURE SHIFT
PRESSURE-SWING ADSORPTION
ACID GAS REMOVAL
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
AMMONIA RECOVERY
O2
SEAL GAS
ASH
CO2 TO ATM
SULPHUR
FUEL GAS
METHANOL
WATER
HP STEAM
AMMONIA
WATER TO DISPOSAL
SOLIDS
HP STEAM TO USERS
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STREAM No	1	2	3	4	5	6	7	8	9	10	11
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP [illegible]% MOISTURE	AS RECEIVED LIGNITE TO STEAM PLANT [illegible]% MOISTURE	LIGNITE TO GASIFIER [illegible]% MOISTURE	OXYGEN TO GASIFIER	BFW TO STEAM PLANT	STEAM TO GASIFIER	ASH FROM STEAM PLANT TO DISPOSAL	ASH FROM GASIFIER TO DISPOSAL	BFW	HP STEAM	QUENCH WATER
FLOW RATE t/h	3835	1412	2307	1464	1155	672	148	377	2078	1977	350

STREAM No	12	13	14	15	16	17	18	19	20	21	22
DESCRIPTION	RAW SYNGAS	QUENCH WATER WITH FINES	SEPARATED WATER TO NH_3 RECOVERY	SLUDGE TO SOLIDS DISPOSAL	NATURAL GAS	ANHYDROUS AMMONIA	STEAM	SOLIDS TO LANDFILL	FUEL GAS	SYNGAS TO METHANOL SYNTHESIS	ACID GAS TO CLAUS PLANT
FLOW RATE t/h	3637	443	848	122	781	6	1677	845	2142	2955	41

STREAM No	23	24	25	26	27	28	29	30
DESCRIPTION	CO_2 TO ATM	FUEL GAS	RAW METHANOL TO PURIFICATION	FUEL GRADE METHANOL	WATER	SULPHUR	SEAL GAS RECYCLE TO GASIFIER	HYDROGEN
FLOW RATE t/h	642	2292	3671	2835	161	38	54	298

33	32	31
WATER TO DISPOSAL	WATER	HP STEAM
480'	14	4504

	DESCRIPTION	DATE
C	GENERAL UPDATE	22.2.84
B	RELEASED FOR FINAL REPORT	9.8.83
A	RELEASED FOR DRAFT REPORT	6.7.83

REVISIONS

REFERENCE DRAWINGS

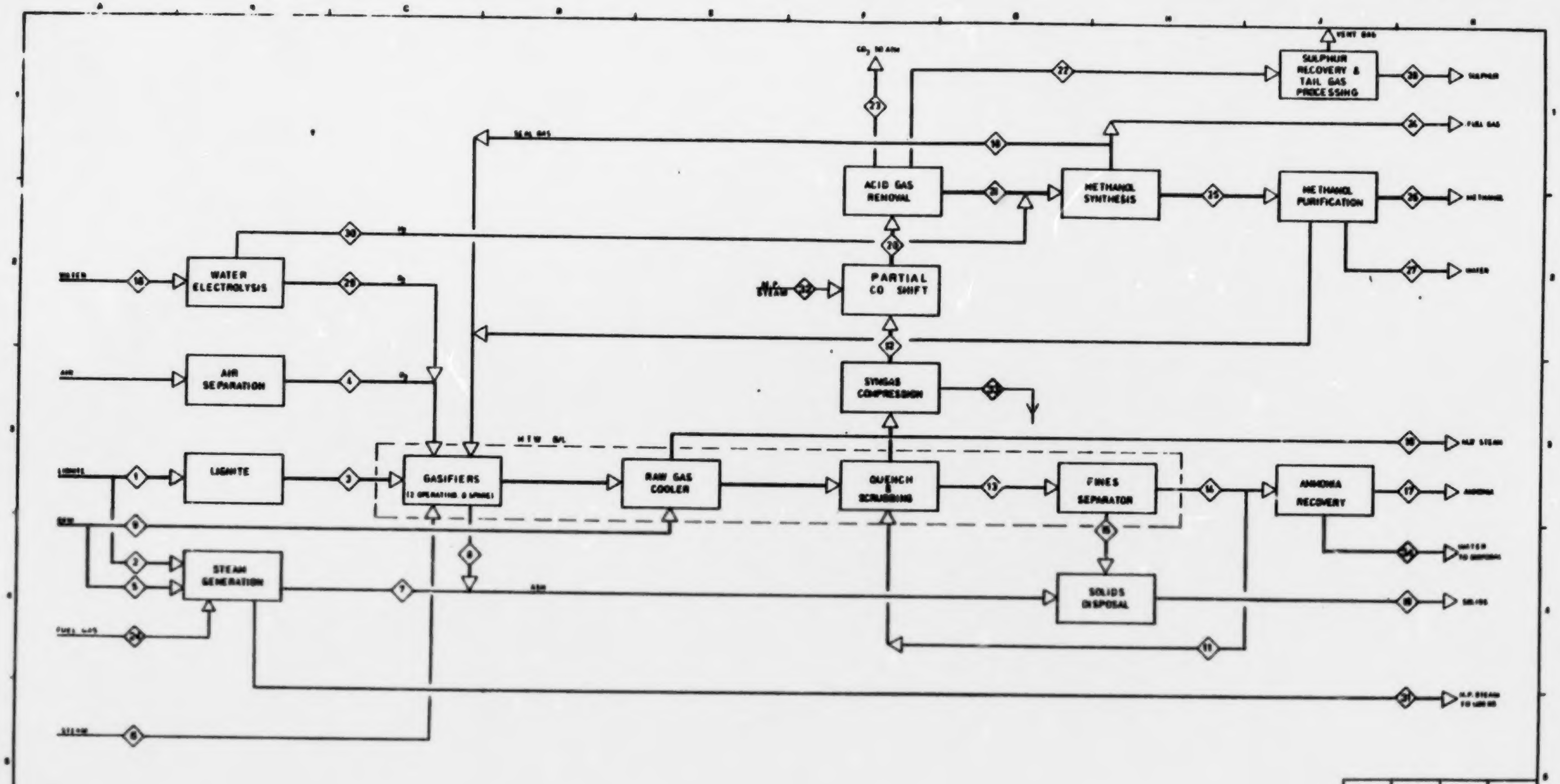
PROCESS
SCALE NONE
DESIGNED AS
DRAWN AS
CHECKED JWC
APPROVED JAM

CLIENT ONTARIO ENERGY CORPORATION
LOCATION TORONTO ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
HIGH TEMPERATURE WINKLER
OPTION 2
STEAM METHANE REFORMING PRESSURE SWING ADSORP

JOB No 3424 16
DRAWING NUMBER 400-43-03 C



WATER
AIR
LIGNITE
BFW
FUEL GAS
STEAM
WATER ELECTROLYSIS
AIR SEPARATION
LIGNITE
STEAM GENERATION
GASIFIERS (2 OPERATING 0 SPARE)
HTW BFL
RAW GAS COOLER
QUENCH & SCRUBBING
FINES SEPARATOR
SOLIDS DISPOSAL
AMMONIA RECOVERY
SYNGAS COMPRESSION
PARTIAL CO SHIFT
M.P. STEAM
ACID GAS REMOVAL
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
SEAL GAS
ASH
CO_2 TO ATM
VENT GAS
SULPHUR
FUEL GAS
METHANOL
WATER
H.P. STEAM
AMMONIA
WATER TO DISPOSAL
SOLIDS
H.P. STEAM TO USERS
H_2
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34

STREAM No	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP (35% MOIST.)	AS RECEIVED LIGNITE TO STEAM PLANT (35% MOIST.)	LIGNITE TO GASIFIERS (12% MOIST.)	OXYGEN TO GASIFIERS	BFW TO STEAM PLANT	STEAM TO GASIFIERS	ASH FROM STEAM PLANT TO DISPOSAL	ASH FROM GASIFIER TO DISPOSAL	BFW	H.P. STEAM	QUENCH WATER	RAW SYNGAS TO CO SHIFT	QUENCH WATER WITH FINES	DEFINED WATER TO NH_3 RECOVERY	SLUDGE TO SOLIDS DISPOSAL	DEMIN WATER	ANHYDROUS AMMONIA
FLOW RATE (t/d)	3835	1343	2387	537	826	569	138	376	2076	1877	380	3837	970	848	122	1086	8

STREAM No	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34
DESCRIPTION	TAIL GAS RECYCLE TO GASIFIER	SOLIDS TO LANDFILL	SHIFTED SYNGAS TO ACID GAS REMOVAL	SYNGAS TO METHANOL SYNTHESIS	ACID GAS TO CLAUS PLANT	CO_2 TO ATM	FUEL GAS	RAW METHANOL TO PURIFICATION	FUEL GRADE METHANOL	WATER	SULPHUR	OXYGEN	HYDROGEN	H. P. STEAM	M. P. STEAM	WATER	WATER TO DISPOSAL
FLOW RATE (t/d)	54	280	4086	2272	41	1753	185	2113	390	103	39	964	121	8363	864	14	480

C	GENERAL UPDATE	21.3.84
B	RELEASED FOR FINAL REPORT	30.6.83
A	RELEASED FOR DRAFT REPORT	8.7.83

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SCALE NONE

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LOCATION TORONTO ONTARIO

KILBORN

TITLE
LIGNITE INDIRECT LIQUEFACTION
HIGH TEMPERATURE WINKLER
OPTION 3
ELECTROLYSIS

3626	16

DRAWING NUMBER
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Ontario Lignite Indirect Liquefaction Study Phase 2 Lignite Testing and Phase 1 Update

Prepared for:
Ontario Energy Corporation

December 1983
Kilborn Limited

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Introduction

1.0 INTRODUCTION

This phase of the study involved testing of Onakawana lignite in the gasification technology developer's facilities to assess its potential to produce a syngas for conversion to methanol. Also involved, although not part of the original scope for Phase 2, was an update of the process and cost data provided in Phase 1, based on the test data and predictions provided by the technology developer.

Phase 1 of the study assessed four gasification technologies combined with various processing steps for the conversion of lignite to fuel grade methanol. This assessment consisted of preliminary process design and cost estimates for a grass-roots plant located in Onakawana, Ontario. A lignite only base case and three co-processing options were considered for each technology. The co-processing options entailed blending the lignite-produced syngas with a syngas produced from natural gas, or hydrogen produced either from natural gas or by water electrolysis.

It was recommended in Phase 1 to pursue the test program with Rheinbraun, the technology developer for the High Temperature Winkler gasification process. The High Temperature Winkler process is a fluidized bed gasification technology and is a derivation of the first generation Winkler process which has been well proven using lignite as a feedstock. Rheinbraun have also extensively used lignite in their High Temperature Winkler bench scale and pilot plant facilities and are presently constructing a demonstration plant to convert Rhenish brown coal to methanol via the High Temperature Winkler process.

Approximately 360 kg of Onakawana lignite sample (45 percent moisture basis) was shipped to Rheinbraun's facilities for testing in the bench-scale unit. The objectives of the test were to verify the behaviour of Onakawana lignite in a fluidized bed gasifier and to provide performance data. This performance data was interpreted by Rheinbraun to predict key process parameters for a commercial size High Temperature Winkler gasifier. The key parameters relevant to this study consist of the following:

- Raw gas composition
- Raw gas yield
- Carbon Conversion
- Steam and oxygen demands for gasification

These data were provided by Rheinbraun to Ontario Energy Corporation following a preliminary review of the test results at their facilities in West Germany.

This report presents the following:

1) Conclusions drawn from the test report issued by Rheinbraun and the subsequent process and cost data generated as a result of the report.

2) Recommendations relevant to further evaluation of lignite to methanol processes utilizing Onakawana lignite as a feedstock.

3) Revised Phase 1 process data for the lignite only base case, the natural gas co-processing and electrolytic hydrogen co-processing options.

4) Revised Phase 1 preliminary estimates for 2500 t/d fuel grade methanol plants via the processing routes described in (3) above. Capital costs, operating costs and total product costs are provided for the base case and two options.

Conclusions and Recommendations

2.0 CONCLUSIONS AND RECOMMENDATIONS

2.1 CONCLUSIONS

2.1.1 The primary conclusion from the Test Report submitted by Rheinbraun is that Ontario lignite is well suited for High Temperature Winkler fluidized bed gasification. The gasification behaviour of the lignite is similar to that of Rhenish brown coal, with which Rheinbraun has developed the High Temperature Winkler process. The main differences in characteristics between Ontario lignite and Rhenish brown coal were in particle size, ash softening point and ash content and these values are compared as follows:

	Ontario Lignite	Rhenish Brown Coal
Average grain diameter (um)	136	463 (Typical)
Ash softening point (^{o}C)	1260	1320 - 1370
Ash content (% wt, mf)	27.9	1.5 - 5

Particle Size

The lower particle size used in the test contributed to the poor carbon conversion experienced due to the excessive carry-over of dust from the gasifier to the cyclones. As only a limited quantity of lignite was available for the test runs, Rheinbraun were unable to control the size reduction of this size of sample. Rheinbraun considered that for a larger size sample or for commercial operation, the particle size of Ontario lignite could be controlled and maintained comparable to the size range used for Rhenish brown coal.

Ash Softening Point

Rheinbraun noted this difference between the Ontario lignite and that typical for Rhenish brown coal, but did not consider it a drawback, particularly for a commercial application.

Ash Content

Rheinbraun reported a significantly higher ash content (27.9 percent) than that given in the typical Ontario lignite analysis (18.8 percent). Even at the lower ash content this is considerably higher than that for Rhenish brown coal. Rheinbraun experienced difficulties in the bench-scale test runs with ash removal but this was due again to the limited size of sample as normally several runs are required to establish the optimum ash removal rate. Rheinbraun have demonstrated in the 35 t/d pilot plant that high ash lignites can be handled effectively, and they do not consider the handling of Ontario lignite ash a problem.

2.1.2 Process Update

The Phase 1 process data for the base case and two co-processing options were revised in this phase of the study to reflect the raw gas composition and yield predicted by Rheinbraun. The process design was also revised to fix the capacity of methanol plant at 2500 t/d for the base case and two options. In Phase 1 the base case and electrolytic hydrogen co-processing designs were 2010 t/d methanol capacity, and the natural gas co-processing design was 2780 t/d capacity.

The main impact on the process design is the lower CO and H_2 content in the gasifier-produced raw gas predicted by Rheinbraun than that assumed in the Phase 1 design. This lower yield has resulted in the following adjustments to the Phase 1 material balance requirements on a per tonne of methanol basis:

	Lignite Feed to gasifiers	Natural Gas Feed	Electrolytic H_2 Feed
Lignite Only	+ 11%	Not applicable	Not applicable
Lignite/Natural Gas	+ 53%	- 8%	Not applicable
Lignite/Electrolytic H_2	No Change	Not applicable	+ 30%

The 30% increase in electrolytic hydrogen for the lignite/electrolytic hydrogen option is only partially due to the lower CO and H_2 content of the lignite derived raw gas. The remainder of the increase results from an optimization of processing units to eliminate an air separation plant (which was included in Phase 1).

The thermal efficiency, as shown on Table 4-2 falls within the range of that reported by the process developer for the lignite only case and, for the natural gas option, the efficiency is superior to lignite only but lower than the commercially-proven efficiency range of 65-67 percent for a conventional natural gas to methanol process.

2.1.3 Cost Update

A summary of the capital and operating costs, and total product costs is presented in Table 2-1. The product costs are given at three Capital Recovery Factors of 10, 15 and 20 percent. At the lowest Capital Recovery Factor of 10 percent which is representative of a public utility investment criterion, the base lignite only case displays a marginally lower product cost than the lignite/natural gas co-processing option, and a significantly lower product cost than the lignite/electrolytic hydrogen co-processing option. However, with increased capital recovery factors of 15 and 20 percent, in the range of a private investment criterion, the natural gas co-processing option appears marginally more favourable than the lignite only case, reflecting the increased capital cost of the lignite only facilities. The lignite/electrolytic hydrogen co-processing option product costs are still significantly higher than the competing processes at the 15 and 20 percent Capital Recovery Factors.

Tables 2-2 and 2-3 depict the effect on product cost by varying the natural gas cost for the lignite/natural gas co-processing option and the power cost for the lignite/electrolytic hydrogen option. A lower natural gas cost of \$.105/*Nm^3 and power cost of 1.8c/kWh were selected. It is evident that in comparing the product costs resulting from the higher gas cost and the lower power cost that the electrolytic hydrogen co-processing option is not competitive with the natural gas co-processing option.

*Nm^3 refers to normal cubic metres, at 0^oC, 760 mmHg.

TABLE 2-1

SUMMARY OF CAPITAL, OPERATING AND PRODUCT COSTS

	FEEDSTOCK COST	LIGNITE ONLY	LIGNITE/NATURAL GAS CO-PROCESSING	LIGNITE/ELECTROLYTIC H_2 CO-PROCESSING
Methanol Capacity, t/y		800 000	800 000	800 000
Capital Cost, $ million		859.9	581.1	786.5
Capital Cost per annual tonne Methanol (MeOH), $		1074.4	726.4	983.1
Feedstock Cost, $/t MeOH				
Lignite	$ 15.00/t	59.62	22.50	35.87
Natural Gas	$ 0.156/Nm^3	-	86.62	-
Power (for Electrolytic H_2)	2.71c/kW hr	-	-	109.75
Other Operating Costs, $		47.37	37.51	35.12
By-Product Credit, $		4.61	1.57	2.82
Total Production Cost, $ (Operating Cost minus By-Product Credit)		102.38	145.06	177.92
Total Product Cost*				
- 10% Capital Recovery		209.82	217.70	276.23
- 15% Capital Recovery		263.54	254.02	325.38
- 20% Capital Recovery		317.26	290.34	374.54

* Total Product Cost: Production Cost + (Capital Cost/t x percent Capital Recovery)

TABLE 2-2

LIGNITE/NATURAL GAS CO-PROCESSING
PRODUCT COSTS VS NATURAL GAS COSTS

	NATURAL GAS COST $\$.156/Nm^3$	NATURAL GAS COST $\$.105/Nm^3$
Methanol Capacity	800 000	800 000
Capital Cost, $ million	581.1	581.1
Capital Cost $/tonne	726.4	726.4
Feedstock Cost $/tonne		
Lignite	22.50	22.50
Natural Gas	86.62	62.50
Other Operating Costs $/tonne	37.51	37.51
By-Product Credit $/tonne	4.57	1.57
Total Production Cost, $/tonne	145.06	120.94
Total Product Cost $/tonne		
at 10% Capital Recovery	217.70	193.58
at 15% Capital Recovery	254.02	229.90
at 20% Capital Recovery	290.34	266.62

TABLE 2-3

LIGNITE/ELECTROLYTIC H_2 CO-PROCESSING

PRODUCT COSTS VS POWER COSTS

	POWER COST 2.71c/kWh	POWER COST 1.8c/kWh
Methanol Capacity	800 000	800 000
Capital Cost, $ million	786.5	786.5
Capital Cost $/tonne	983.1	983.1
Feedstock Cost $/tonne		
Lignite	35.87	35.87
Power	109.75	72.87
Other Operating Costs, $/tonne	35.12	35.12
By-Product Credit, $/tonne	2.82	2.82
Total Production Cost, $/tonne	177.92	141.04
Total Product Cost $/tonne		
at 10% Capital Recovery	276.23	239.35
at 15% Capital Recovery	325.38	288.50
at 20% Capital Recovery	374.54	337.66

2.2 RECOMMENDATIONS

The product cost data presented in Section 5.0 of this report are being utilized by Ontario Energy Corporation as part of a separate study to assess the most financially attractive methanol process from a variety of feedstocks. Lignite feedstock and the two lignite co-processing options have been evaluated in this study. A recommendation on which option to adopt is therefore outside the scope of this study. However, in the event of a more detailed technical and financial assessment of an Ontario lignite to methanol plant the following recommendations should be considered:

It is recommended that the High Temperature Winkler gasification process be utilized as a component of the overall lignite to methanol process for the following reasons:

- The bench scale tests, carried out by Rheinbraun the process developer, indicated that Ontario lignite is well suited to High Temperature Winkler fluidized bed gasification.

- The gasifier-produced raw gas composition and yield predicted by Rheinbraun for a commercial plant do not significantly affect the overall process design and final costs developed in Phase 1 and updated in Phase 2 of this study.

- It appears very likely that the High Temperature Winkler Process will be commercially proven by the study requirement of 1990. A demonstration plant utilizing the High Temperature Winkler Process is planned to start up in West Germany in late 1984.

It is also recommended that the following items be evaluated:

Alternative Plant Location

The plant location, designated within the scope of this study is mine-mouth in Onakawana, Ontario. Onakawana is a remote location, accessible only by rail, and is not well suited for construction and operation of a complex lignite to methanol processing plant. It may be preferable to locate the plant closer to the marketplace, such as in a serviced industrial area in Northern or Southern Ontario, where construction and operating costs are likely to be considerably lower. This would entail shipping the lignite, probably dried to 12-15 percent moisture, to the plant location. This cost is offset by the cost of shipment of 2500 t/d methanol from Onakawana to the marketplace. This alternative is most attractive for the lignite/natural gas co-processing option as the weight of lignite (12% moisture content) consumed within the plant and methanol produced are almost equivalent.

Lignite Sampling and Analysis

Two key lignite parameters, moisture and ash content, which have significant impact on the process selection and subsequent costs, were selected at the outset of this study. A moisture content of 45 percent and an ash content of 18.8 percent (mf basis) were selected as a result of analyses of grab samples of lignite from one location within the overall Onakawana deposit. An ash analysis by Rheinbraun on the lignite sample sent for the gasification tests revealed an average ash content of 27.9 percent (mf basis). Previous analyses performed on Ontario lignite reported moisture contents in the range of 39.7 to 55.2 percent. A comprehensive sampling and statistical analysis program should be performed on the Onakawana lignite deposit prior to detailed process and cost evaluation of lignite to methanol processes. This program would provide a higher level of confidence in the critical parameters for a lignite to methanol process design.

Test Program and Results

3.0 TEST REPORT SUMMARY

3.1 General

Lignite samples, stored in two 200 litre drums, were shipped in July 1983 for testing in Rheinbraun's bench-scale test facilities in Aachen, West Germany. The objective of the test program was to predict the behaviour of Ontario lignite in a commercial-size High Temperature Winkler unit and to obtain key process data so that the Phase 1 conceptual design could be updated. A sample of Saskatchewan lignite was also tested concurrently with the Ontario lignite. Physical, chemical and petrographic analyses of Ontario lignite, which formed the basis for Phase 1 conceptual design, were also provided to Rheinbraun, although these analyses were not of the sample shipped. Rheinbraun repeated some of these tests as part of the test program and one major discrepancy was in ash content, which Rheinbraun reported at 27.9 percent mf versus an average of 18.8 percent mf in the analyses provided. A grab sample of the lignite sent to Rheinbraun was rechecked for ash content by a Canadian laboratory and an average content of 15.5 percent mf was reported.

3.2 Test Program

The test program consisted of three main components:

(1) Physical and chemical tests comprising reactivity, porosity and ignition temperature measurements and screen analysis. The data from these tests supplemented those provided to Rheinbraun by Ontario Energy Corporation.

(2) Cold flow fluidized bed tests on the crushed and dried lignite in the 200 mm diameter glass model.

(3) Steady state, gasification tests in the bench-scale unit (PDU).

3.3 Test Report

As the test report has already been provided to Ontario Energy Corporation and members of the Advisory Committee, it is not duplicated in this report. The test results on the Saskatchewan lignite are also contained in the test report.

3.4 Gasifier Performance Predictions

Rheinbraun provided, separate from the test report, a comparison of gasification data between Ontario lignite and Rhenish brown coal for bench-scale (PDU), pilot plant and a commercial-size plant. The predicted data for the performance of Ontario lignite in a commercial size gasifier was used for the Phase 1 update described in Section 4.0.

Phase 1 Process Update

4.0 PHASE 1 PROCESS UPDATE

4.1 General

The conceptual design of lignite to methanol plants of varying capacities between 2000 and 3000 t/d, incorporating four gasification technologies, was performed in Phase 1 of this study. In this section the relevant Phase 1 conceptual design is revised to reflect the predictions made by Rheinbraun, the High Temperature Winkler process developer. The design has been revised for the base case, Option 1 (Lignite + Natural Gas) and Option 3 (Lignite + Electrolytic Hydrogen) of the High Temperature Winkler technology described in Section 3.5 of the Phase 1 report.

The highlights of the revisions are as follows:

- All three designs are now based on a 2500 t/d capacity fuel grade methanol plant.
- All three designs are modified to reflect the gasification agents consumption, syngas composition, syngas yield and carbon conversion predicted by Rheinbraun for a commercial size plant with High Temperature Winkler gasifier(s).
- All three designs are modified to reflect the Rheinbraun recommendation that the gasifier lockhopper seal gas be changed from purge gas (used in Phase 1) to carbon dioxide.
- The electrolytic hydrogen Option 3 design in Phase 1 has been modified to exclude an air separation plant. Thus the total oxygen requirement for the gasifier will be supplied from the electrolysis plant.

The facilities descriptions for the Base Case and options were provided in detail in the Phase 1 report. Hence these descriptions will not be included in this report. However, there have been changes to facilities, notably to the methanol synthesis process and utilities, which are described in sections 4.2, 4.3 and 4.4 to follow. Block flow diagrams for the Base Case and Options are presented as follows:

Base Case

Drawing 400-43-05 describes a lignite to methanol plant with all the carbon requirements provided by the lignite, and the H_2/CO ratio being adjusted by CO shift conversion.

Natural Gas Co-Processing Option

Drawing 400-43-06 describes a co-processing scenario of lignite and natural gas. Natural gas is delivered to the plant via a dedicated new pipeline from the Trans-Canada Pipeline. The natural gas is steam reformed to produce a hydrogen-rich syngas which is combined with the gasifier produced syngas for methanol synthesis.

Electrolytic Hydrogen Option

Drawing 100-43-07 describes a lignite to methanol plant in which electrolytic hydrogen is used to supplement the hydrogen-deficient gasifier produced syngas. Oxygen produced in the electrolysis plant is used in the gasifiers.

4.2 Process Design Basis

A summary of the key criteria for the Phase 2 design update is as follows:

4.2.1 Lignite

Onakawana lignite analysis, as used for the Phase 1 design, with the following composition:

Proximate Analysis, wt%

Moisture	45.0
Ash, mf	18.81
Volatile Matter, mf	40.49
Fixed Carbon, mf	40.70
Heating Value, kJ/kg, mf	22348

Ultimate Analysis, wt% db

Carbon	59.30
Hydrogen	3.81
Oxygen	15.97
Nitrogen	.70
Sulphur	1.41
Ash	18.81

Rheinbraun performed Proximate and Ultimate analyses as part of the test program and reported an ash content of 27.9 percent, mf. A grab sample of the lignite sent to Rheinbraun was rechecked for ash content by a Canadian laboratory and an average content of 15.5 percent mf was reported. The original value of 18.8 percent mf was used for this design update.

4.2.2 Gasifiers

High Temperature Winkler design

Oxygen consumption	0.51 Nm^3/kg lignite, maf
Steam consumption	0.30 kg/kg lignite, maf
CO_2 recycle	0.125 Nm^3/kg lignite, maf
Lignite feed moisture content:	12%
Carbon conversion:	95%
Specific Synthesis Gas yield:	1655 Nm^3 ($CO + H_2$)/t lignite, maf

		Vol %
Raw gas analysis,	H_2	33.0
H_2O and	CO	44.5
N_2 free	CO_2	19.9
	CH_4	2.0
(Stream 12 on	H_2S	0.6
drawings 100-43-05,		
06, 07)		
		100.0

4.2.3 Steam-Methane Reforming

Reformed Gas Analysis:		Vol %
	H_2	76.2
	CO	16.9
	CO_2	6.9
		100.0

4.2.4 Methanol Synthesis and Purification

I.C.I. Process - Single Train
Capacity: 2500 t/d Fuel Grade Methanol

4.3 Process Facilities

4.3.1 Lignite Preparation

Lignite drying is now carried out using rotary steam tube dryers exclusively as opposed to two-stage steam and flue gas drying considered in Phase 1. Rheinbraun have demonstrated that Rhenish brown coal can be successfully dried in steam tube dryers from 60 percent moisture content to the 12 percent required for gasification.

4.3.2 Gasification

The gasifier sizing and sparing has been based on a maximum capacity of 1200 t/d of lignite feed (12 percent moisture) per gasifier and one spare for a maximum of four operating gasifiers. Gasifiers have been downsized as required to concur with this philosophy. A summary of the gasifier requirements is as follows:

	Base Case	Natural Gas Option	Electrolytic H_2 Option
Lignite feed (t/d)	4842	1644	2995
No. of operating gasifiers	4	2	3
Spares	1	1	1
Gasifier size (percent of max. available)	100	68.5	83

Carbon dioxide is used as a seal gas in the bottom cone of each gasifier to protect the ash removal screw conveyor. Purge gas was assumed for the Phase 1 design, but Rheinbraun consider CO_2 a better and safer seal medium, as some leakage can occur and also the CO_2 is readily available.

4.3.3 Shift Conversion

The initial objective of the process design update was to eliminate the shift conversion unit for the electrolytic hydrogen option. However, by eliminating this unit and providing all the hydrogen necessary to adjust the H_2/CO ratio with hydrogen from the water electrolysis plant, an oxygen excess of 414 t/d (over that required for the gasifiers) resulted. Accordingly, the design has been optimized by increasing the lignite feed by about 10 percent, shifting about 10 percent by weight of the CO in the syngas and reducing the oxygen vent to essentially zero.

4.3.4 Methanol Synthesis and Purification

For the Phase 1 design, a methanol synthesis unit based on the Lurgi low pressure process was proposed. For Phase 2 the ICI low-pressure methanol process is considered, the prime reason being that the Lurgi process has not been proven at 2500 t/d methanol production on a single train basis. The ICI process is similar to the Lurgi process in process configuration with the following major differences in the proposed design:

- Synthesis pressure is 8000 kPa rather than 5500 kPa for the Phase 1 design
- The synthesis reaction in the ICI methanol converter is quenched by injecting cool gas through specially designed spargers, and also by generating steam in external heat exchangers. The Lurgi methanol converter is a tubular reactor with medium pressure steam being generated in the shell.

4.4 Utilities

4.4.1 General

Phase 1 utility consumptions, comprising steam, fuel gas, raw water, cooling water, boiler feed water and power, have been adjusted as required for the Phase 2 design bases. Power is assumed to be purchased from the provincial grid for all cases as described in the Phase 1 report with the existing line upgraded for the electrolytic hydrogen option. The most significant change to utility distributions from the Phase 1 design has been to the steam and fuel gas balances, as described in more detail below. The utility requirements for the three processing options are shown on Table 4.1.

4.4.2 Steam/Fuel Gas

Energy balances have been changed from Phase 1 to reflect the following:

- Lignite drying performed exclusively by low pressure (0.55 MPa) steam.
- Methanol production has changed to 2500 t/d for the Base Case and Options.
- The purge gas calorific value and quantity available for heat recovery is greater than that used for Phase 1 design. The purge gas is now used as boiler fuel for the lignite only base case and lignite/electrolytic hydrogen option and as reformer fuel for the lignite/natural gas option.
- Syngas flow has increased even greater than proportional to the methanol production change due the lower H_2 and CO yield predicted by Rheinbraun.

Steam production and consumptions for the three cases are presented on steam distribution diagrams, drawings B400-43-08, B400-43-09, B400-43-10. Steam is utilized in the complex at three pressure levels, 10.8 MPa, 3.5 MPa and 0.55 MPa.

10.8 MPa steam

Steam at 10.8 MPa, superheated to 480°C, is generated as follows: Lignite (as-received) and fuel gas fired boilers provide the steam for the lignite only base case and lignite/electrolytic hydrogen cases. For the lignite/natural gas co-processing option a lignite fired boiler and the steam methane reforming furnace provide the 10.8 MPa superheated steam. The thermal efficiency of the lignite fired boiler is 75 percent and the fuel gas fired boiler is 80 percent. This high pressure superheated steam is used exclusively for backpressure or condensing steam turbine drivers for the large compressors. The backpressure turbines exhaust steam to the 3.5 MPa and 0.55 MPa levels as shown on the distribution diagrams.

3.5 MPa

Medium Pressure saturated steam at 3.5 MPa is generated in the gasifier waste heat boiler and this steam, with the steam exhausted from the backpressure turbines provides the steam requirements for the gasifiers, CO shift, acid gas removal, sulphur recovery and ammonia recovery units.

0.55 MPa

Low Pressure saturated steam is provided from the backpressure turbine exhaust, CO shift unit and methanol synthesis unit. The users are the lignite drying plant, methanol purification unit and miscellaneous low pressure users such as heat tracing, building heating and utility stations.

4.5 Thermal Efficiency

The overall thermal efficiencies based on higher heating value of the feedstock and products are given on Table 4-2. Electrical power energy equivalent is taken at the 100 percent efficiency value of 3.6 GJ/Mw hr for the base case and natural gas co-processing option. For the electrolytic hydrogen option the thermal efficiencies are given for two values of power energy, the 100 percent value as above and a value representing 34 percent efficiency (10.5 GJ/MWh).

TABLE 4-1

2500 T/D METHANOL PLANTS

UTILITY REQUIREMENTS

PROCESSING OPTION UTILITY	LIGNITE ONLY	LIGNITE/NATURAL GAS CO-PROCESSING	LIGNITE/ELECTROLYTIC H_2 CO-PROCESSING
Steam Generation, t/d	10337	3578	6525
Cooling Water Circulation, t/d	536000	308000	351000
Demineralized Water, t/d	1324	450	821
Raw Water, t/d	27000	15200	17700
Power, mW	17.4	14.3	422

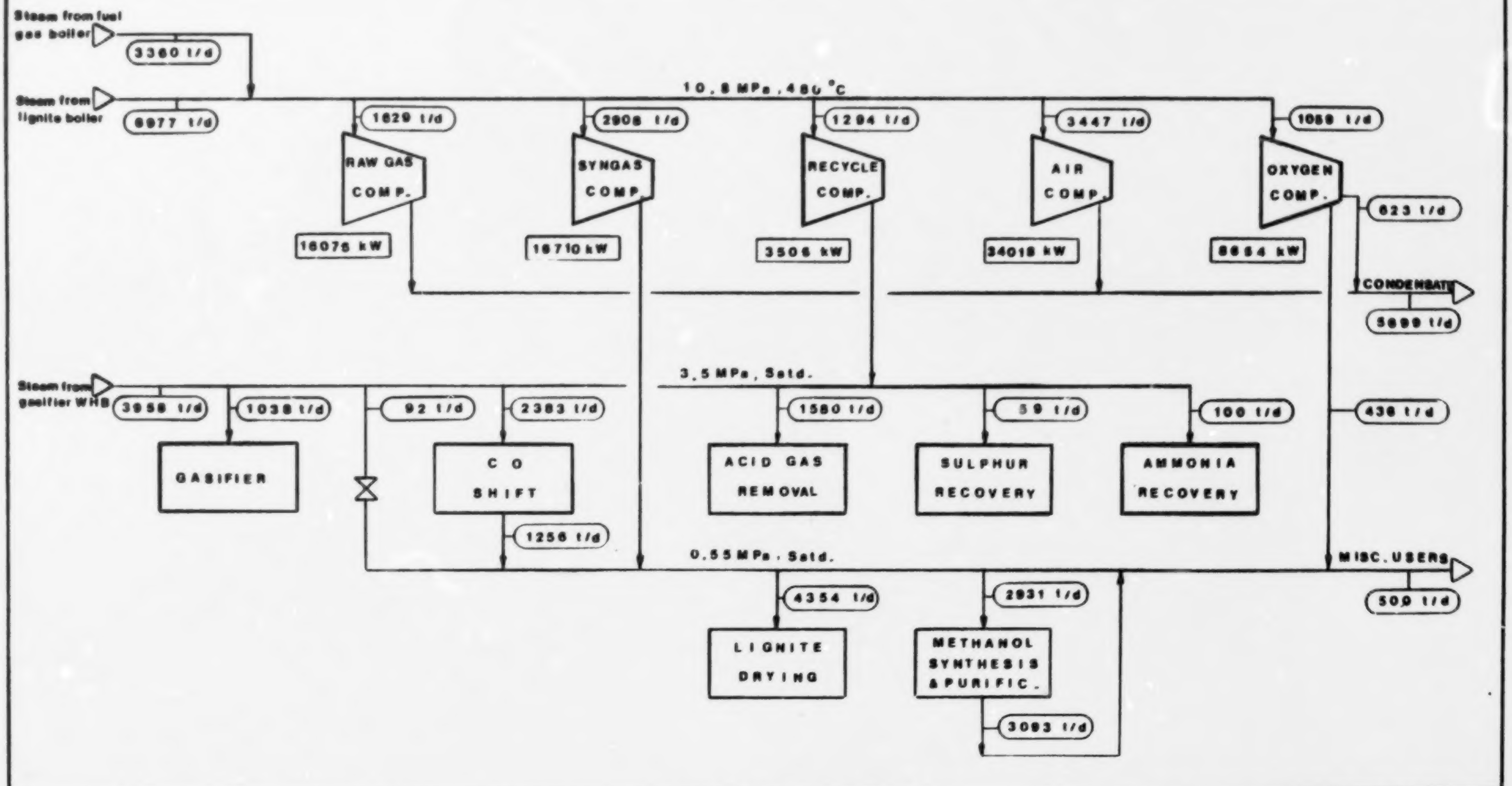
TABLE 4-2

THERMAL EFFICIENCIES

FEED:	H.H.V. GJ/t	LIGNITE ONLY		LIGNITE/NATURAL GAS		LIGNITE/ELECTROLYTIC H_2	
		Quantity	GJ/d	Quantity	GJ/d	Quantity	GJ/d
Lignite to Gasifiers, (45% m.c.), t/d	12.282	7748	95161	2630	32302	4792	58855
Lignite to Utilities, (45% m.c.), t/d	12.282	2185	26836	1120	13756	1194	14665
Natural Gas to Process, t/d	52.825	-	-	727	38388	-	
Natural Gas to Fuel, t/d	52.825	-	-	143	7575	-	
Power at 100% Efficiency MW		17.4	1503	14.3	1236	422	36461
Power at 34% Efficiency MW						422	106850
Total Energy in Feed			123500		93257		109981* 180370**
PRODUCTS:							
Fuel Grade Methanol t/d	22.7	2500	56750	2500	56750	2500	56750
Sulphur t/d	9.264	60	556	20	185	37	343
Ammonia t/d	18.61	23	428	8	149	14	261
Total Energy in Products			57734		57084		57354
Thermal Efficiency (%) = (Energy in Products) / (Energy in Feed) x 100			46.7*		61.2*		52.1* 31.8**

* Power at 100 percent efficiency

** Power at 34 percent efficiency (10.5 GJ/MWh)


Steam from fuel gas boiler
3360 t/d
Steam from lignite boiler
6977 t/d
10.8 MPa, 480 °C
1629 t/d
RAW GAS COMP.
16075 kW
2908 t/d
SYNGAS COMP.
18710 kW
1294 t/d
RECYCLE COMP.
3506 kW
3447 t/d
AIR COMP.
34018 kW
1059 t/d
OXYGEN COMP.
8654 kW
623 t/d
CONDENSATE
5699 t/d
Steam from gasifier WHB
3958 t/d
1038 t/d
GASIFIER
92 t/d
2383 t/d
C O SHIFT
1256 t/d
3.5 MPa, Satd.
1580 t/d
ACID GAS REMOVAL
59 t/d
SULPHUR RECOVERY
100 t/d
AMMONIA RECOVERY
436 t/d
0.55 MPa, Satd.
4354 t/d
LIGNITE DRYING
2931 t/d
METHANOL SYNTHESIS & PURIFIC.
3093 t/d
MISC. USERS
500 t/d

REFERENCE DWG S	DRAWING No	DESCRIPTION

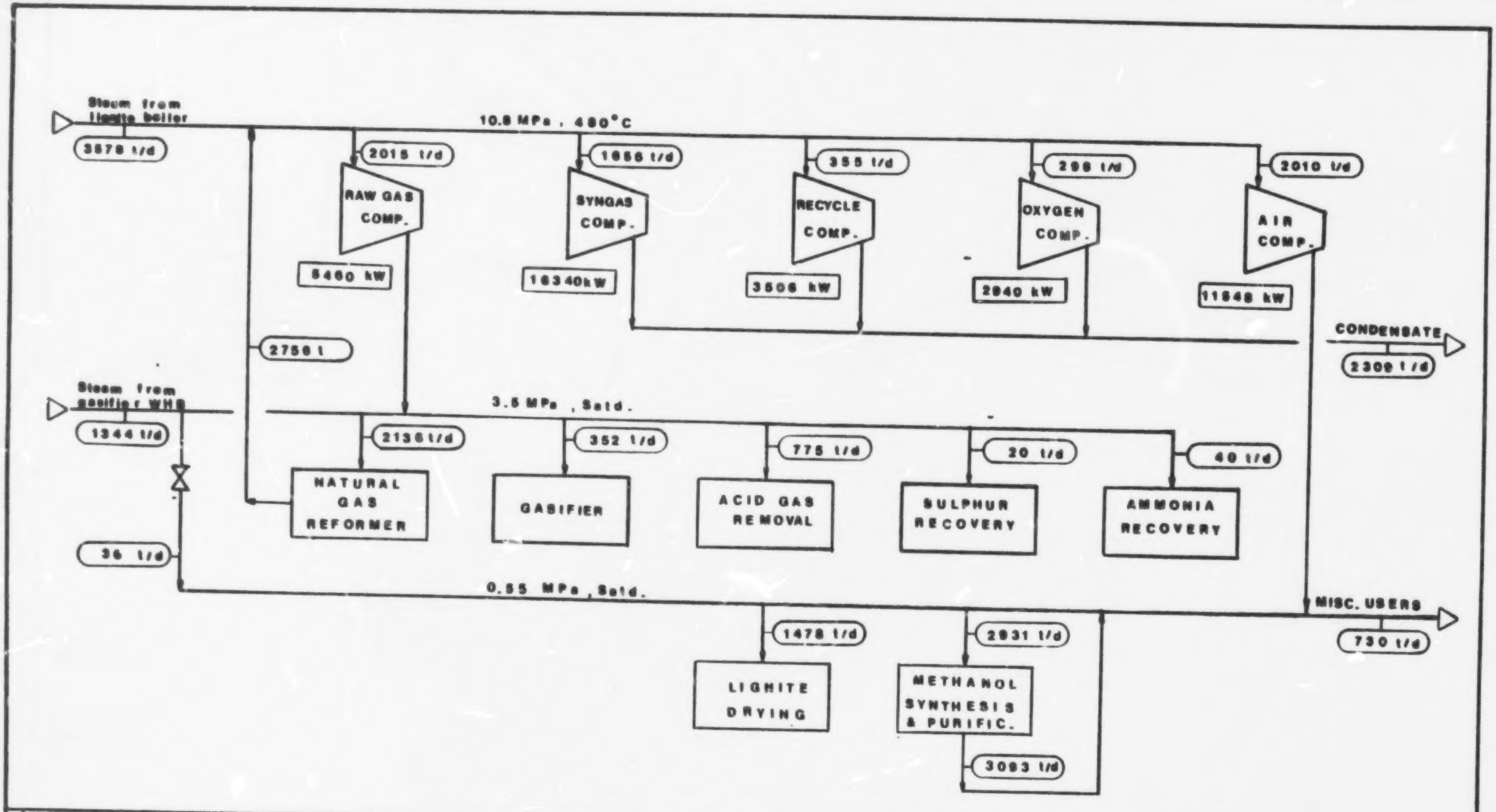
No	DESCRIPTION	DATE	BY				
A	Released for phase 2A final report.	22/1 1984	ML			JM	

CLIENT O. E. C.
LOCATION TORONTO, ONT.

STEAM DISTRIBUTION DIAGRAM
LIGNITE ONLY

KILBORN

SECTION	Process		PROJECT No	DIVISION No
SCALE	—	DATE	3424	17
DRAWN	M.L.		DRAWING NUMBER	REV
CHECKED	H.E.A.		B 400-43-08	A
APPROVED	J.Mc.M.		SHEET OF	


Steam from lignite boiler
3578 t/d
10.8 MPa, 480°C
2015 t/d
RAW GAS COMP.
5460 kW
1656 t/d
SYNGAS COMP.
16340kW
355 t/d
RECYCLE COMP.
3506 kW
298 t/d
OXYGEN COMP.
2940 kW
2010 t/d
AIR COMP.
11848 kW
CONDENSATE
2309 t/d
2756 t
Steam from gasifier WHB
1344 t/d
3.5 MPa, Satd.
2136 t/d
NATURAL GAS REFORMER
352 t/d
GASIFIER
775 t/d
ACID GAS REMOVAL
20 t/d
SULPHUR RECOVERY
40 t/d
AMMONIA RECOVERY
36 t/d
0.55 MPa, Satd.
1478 t/d
LIGNITE DRYING
2931 t/d
METHANOL SYNTHESIS & PURIFIC.
3093 t/d
MISC. USERS
730 t/d

REFERENCE DWG S	
DRAWING No	DESCRIPTION

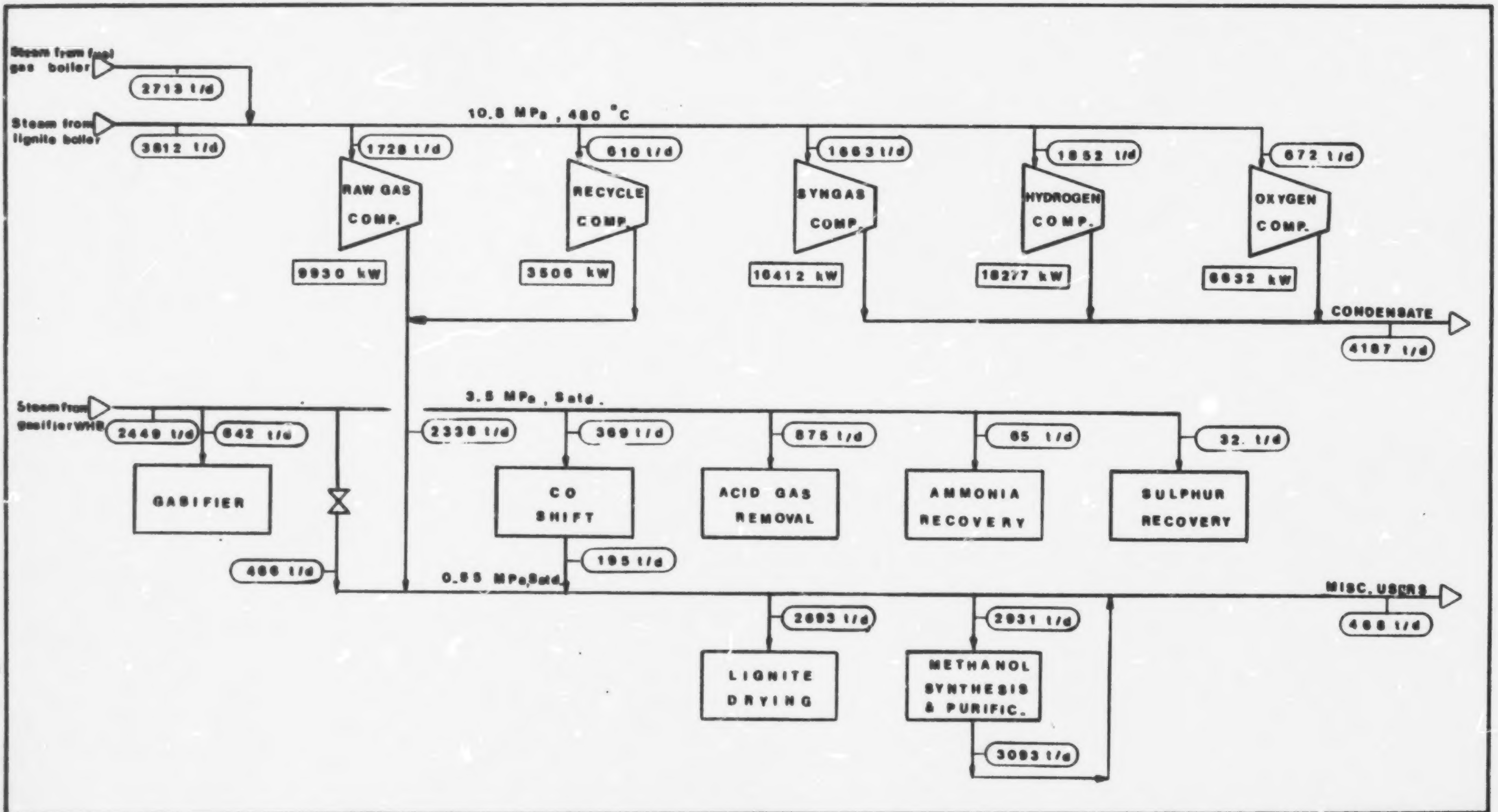
REVISION				
A	Released for phase 2A final report.	22/1 1984	ML	JM
No	DESCRIPTION	DATE	BY	

CLIENT O. E. C.
LOCATION TORONTO, ONT

STEAM DISTRIBUTION DIAGRAM
LIGNITE/NATURAL GAS
CO-PROCESSING

KILBORN

SECTION	Process		PROJECT No	DIVISION No
SCALE	—	DATE	3424	17
DRAWN	M.L.		DRAWING NUMBER	
CHECKED	N.E.A.		B400-43-09	REV A
APPROVED	C.M.		SHEET	


Steam from fuel gas boiler
2713 t/d
Steam from lignite boiler
3812 t/d
10.8 MPa , 480 °C
1728 t/d
RAW GAS COMP.
9930 kW
610 t/d
RECYCLE COMP.
3506 kW
1663 t/d
SYNGAS COMP.
10412 kW
1852 t/d
HYDROGEN COMP.
18277 kW
672 t/d
OXYGEN COMP.
6632 kW
CONDENSATE
4187 t/d
3.5 MPa , Satd.
Steam from gasifier WHB
2449 t/d
642 t/d
GASIFIER
2338 t/d
369 t/d
CO SHIFT
195 t/d
875 t/d
ACID GAS REMOVAL
65 t/d
AMMONIA RECOVERY
32. t/d
SULPHUR RECOVERY
466 t/d
0.55 MPa,Satd.
2693 t/d
LIGNITE DRYING
2931 t/d
METHANOL SYNTHESIS & PURIFIC.
3093 t/d
MISC. USERS
468 t/d

REFERENCE DWG'S	
DRAWING No	DESCRIPTION

REVISION				
A	Released for phase 2A final report.	22/1 1984	M.L	JM
No	DESCRIPTION	DATE	BY	

CLIENT O.E.C.
LOCATION TORONTO, ONT.

STEAM DISTRIBUTION DIAGRAM
LIGNITE/ELECTROLYTIC H_2
CO-PROCESSING

KILBORN

SECTION Process
SCALE
DRAWN M.L.
CHECKED N.E.A.
APPROVED J.Mc.M.
DATE
PROJECT No 3424
DIVISION No 17
DRAWING NUMBER B400-43-10
REV A
SHEET OF

Phase 1 Cost Update

5.0 PHASE 1 COST UPDATE

5.1 General

This section comprises the update of the cost estimate developed in Phase 1, specific to the High Temperature Winkler gasification technology. The Phase 1 cost data has been revised to reflect the process modifications resulting from the data supplied by Rheinbraun and standardizing on 2500 t/d methanol production for the three cases under consideration.

Preliminary capital, operating and production cost estimates have been prepared for the base case and options of an Onakawana lignite-to-methanol plant. The capital and operating costs have been prepared in first quarter 1983 Canadian dollars for a Northern Ontario location. Operating costs include raw materials, utilities, manpower and maintenance requirements, municipal taxes and insurance. Onakawana plant gate product costs have been prepared taking into account operating costs, by-product credits and capital recovery factors. The sensitivity of production cost to capital recovery factors of 10, 15 and 20 percent is presented.

The capital and operating cost criteria are identical to those presented in Section 4.0 of the Phase 1 report with the exception of steam generation and cooling water systems direct costs which have been increased 20 and 40 percent respectively from Phase 1 costs to reflect new information applicable to these facilities. These Phase 1 criteria are again described in paragraph 5.2 below.

5.2 Capital and Operating Cost Criteria

The primary intent of Phase 2 cost update is to provide estimated capital and operating costs for a lignite only base case and two lignite co-processing options. These costs will be interpreted by Ontario Energy Corporation to make financial comparisons between the base case and options and for further economic comparisons. It must be cautioned that the costs provided in this phase of the study do not include certain off-property costs that will require estimation in more detailed economic viability assessments. These include:

- Possible existing Ontario Northland Railway upgrading to allow shipment of heavy process and construction equipment and transportation of the methanol product to proposed markets.

- Road and rail construction outside plant battery limits.

- Methanol distribution system costs for tank cars, terminals and loading facilities.

- General infrastructure costs.

5.2.1 Assumptions

The following basic assumptions were made for the generation of the capital and operating cost estimates:

i) The plant is a grassroots facility, located at the lignite mine-mouth in Onakawana, Ontario. Lignite is delivered to raw material storage by conveyor. Rail transportation and rail off-loading facilities have not been considered in the estimate.

ii) Electricity is available from the existing grid in the local area to supply up to 50 MW of plant power consumption. This satisfies the purchased power requirements for the lignite only and lignite/natural gas co-processing evaluations. The lignite/electrolytic hydrogen co-processing option entails a maximum load of 422 MW to be purchased from Ontario Hydro. This would require installation of a new power line approximately 48 km long at a cost of $195 000/km. The power line would tie into the existing grid from the Otter Rapids generating station and would be routed north to the Onakawana site. The cost of this installation would be to the project. Ontario Hydro verbally confirmed this information.

iii) Sufficient raw water for plant operation is available from a river adjacent to the plant at no cost to the Owner.

iv) The lignite/natural gas co-processing option entails construction of a new natural gas pipeline to the plant gate from the Trans Canada Pipeline Limited (TCPL) pipeline. Inquiries to Northern and Central Gas, the operator in the area, confirmed that a 12 inch dia. pipeline approximately 180 km long would be required to supply 850 000 Nm^3/d. The line would tie into the existing TCPL pipeline at Smooth Rock Falls, Ontario and would be installed parallel to the existing highway and railroad northward to Onakawana. The installation cost of the pipeline would not be capitalized, but rather included as a demand charge with the operating rate. Thus the installation cost is reflected in the operating cost.

v) Due to the remoteness of the location, a construction camp is provided to house construction and project personnel throughout the construction period. This cost is estimated to be 9.5 percent of the total direct cost of the plant.

vi) Operating labour costs are based on the average rate predominant in Northern Ontario mining locations. No allowance has been made for townsite or travelling costs.

vii) Normal construction techniques for civil work such as excavation, backfill, foundations by spread footing construction, etc. has been assumed. No allowance has been made for importation of backfill and special foundation construction such as piling.

5.2.2 Capital Costs

Capital costs were derived using Kilborn in-house estimating information and data from process developers and licensors. Where cost data were not specific to a Northern Ontario location, they were converted to a cost in U.S. dollars, Gulf Coast location. This cost was then converted from U.S. Gulf Coast to a Northern Ontario location using a factor: Canadian Dollars, Northern Ontario location, = 1.5 x U.S. dollars, Gulf Coast location.

This conversion factor is made up of two components - a materials component representing 60% of the value of the plant and a labour component representing 40% of the plant cost.

For the materials component, a factor of 1.07 is used to represent 18% duty on the imported material (about 40% of total material). An exchange factor of 1.25 is used to convert U.S. dollars to Canadian, and a factor of 1.05 is used to represent 5% freight allowance on all material. This results in a materials factor of:

$$0.6\ (1.07 \times 1.25 \times 1.05) = 0.843$$

The labour factor is comprised of the ratio of the base labour rate for a representative trade, such as pipefitting, in Canada to that in the Gulf Coast, each expressed in its own currency. This is multiplied by a productivity factor for Northern Ontario compared to the Gulf Coast, estimated to be 1.2. The labour factor is then:

$$0.4 \left(\frac{18.00}{14.81} \times 1.2 \right) = 0.583$$

A winterization factor of 1.05 is applied to the combined total to reflect the northern location, necessitating extra insulation and heat tracing on plant piping, equipment and instrumentation, to give a total conversion factor:

$$(0.843 + 0.583)\ 1.05 = 1.497, \text{ use } 1.5.$$

Buildings and site development costs were factored from direct costs. Building costs were assessed as two percent of the process unit costs and site development as five percent of the total direct costs.

Costs are presented for the base case and two options in Canadian dollars, Northern Ontario location. The costs are subtotalled for on-site process units and offsites, utilities and services. Site development and building costs are then added, resulting in Total Direct Cost for the plant. A total of 20 percent of direct costs is added to represent home office engineering and indirect field costs such as tools, equipment rental and contractors fees. A contingency of 14 percent is applied to the total direct costs by taking into account complexity and the degree of confidence in the direct cost estimate for the individual process units. Total Installed Cost is computed by adding the Total Direct Costs to the Indirect Cost.

As discussed previously a new power transmission line is included in the lignite/electrolytic hydrogen option. This is assessed at $10.8 million and applied accordingly.

Owners costs, which include Owner's engineering and project management costs, an Initial Catalyst and Chemicals Cost and the Construction Camp cost are added to the Total Installed Cost to give a Total Estimated Cost.

5.2.3 Operating Costs

Operating costs consist of the costs of raw materials, purchased utilities and services, labour, maintenance and overheads. Annual operating costs are derived consistent with an operating factor of 320 days operation per year.

Operating costs comprise the following:

.1 Raw Materials

Lignite

Lignite is assumed to be received at the plant gate, minus 50 mm size and containing 45% moisture. The cost of lignite supplied to the plant is $15.00/t. This cost has been confirmed by Manalta Coal as being the minimum acceptable cost.

Natural Gas

A cost of $0.156/Nm^3 has been estimated to be representative for the study. A pipeline demand charge of $8 236 000 per year to cover pipeline construction, maintenance and operating costs is also levied. This charge is based on an annual service of

265 000 000 Nm^3, and is factored accordingly to reflect the actual natural gas volume relevant to this study. These costs have been estimated from preliminary information provided by the operating gas company, Northern and Central Gas.

.2 Labour

The labour component comprises the following categories:

a) Operating Labour. The operating manpower content is estimated to be 114 for the base case and two options. Labour cost is averaged at $30,000/yr, exclusive of payroll burdens which are described under (e) below.

b) Plant overhead costs, which are estimated at 30% of operating labour.

c) Local administration, to cover plant management and supervision, estimated at 20% of operating labour.

d) Corporate overhead costs, which include head office and sales overhead allocation, estimated at 10% of operating labour.

e) Burdens and supplies, totalling 30% of operating labour consisting of 26% for payroll burdens, and 4% for office equipment, supplies etc.

.3 Insurance

Insurance is estimated at 0.5% of Total Installed Cost of the facilities.

.4 Taxes

An allowance for taxes estimated at 0.25% of Total Installed Cost represents possible municipal taxes applicable to the plant facilities.

.5 Maintenance

Maintenance costs are split into two categories, labour and materials as follows:

Labour (approximately 90 persons): 1.5 x Operating Labour

Materials (Spare Parts, etc.): 1% of Total Installed Cost for Base Case and Lignite/Natural Gas Co-processing option.

: 1% of (Total Installed Cost minus cost of Electrolysis facilities) for the lignite/electrolytic Hydrogen option to reflect the low maintenance electrolysis plant.

.6 Catalyst and Chemicals

2% of Process Unit Direct Costs.

.7 Utilities

Power - Ontario Hydro provided a rate of 2.71 cents per kWh which applies for consumptions up to 500 MW.

Lignite - Lignite to the steam boilers is costed at the same rate as run-of-mine lignite, $15.00/t.

5.2.4 Production Costs

This cost is computed by subtracting the by-product credits from the Total Operating Cost.

.1 By-product Credits

Saleable by-products are only considered for credit. Solid sulphur and anhydrous ammonia have been assessed the following credits:

Sulphur:	\$100/t
Anhydrous Ammonia:	\$240/t

5.2.5 Total Product Cost

Total Product Cost has been presented for three Capital Recovery Factors, 10 percent/yr, 15 percent/yr and 20 percent/yr. The Capital Recovery Cost in \$/tonne is computed as follows:

$$\frac{\text{Total Estimated Costs (\$)}}{\text{Capacity of Plant (tonne/yr)}} \times \frac{\text{Percent Capital Recovery}}{\text{Yr}}$$

The total of the Production Cost and the Capital Recovery Cost equals the Total Product Cost.

5.3 Estimates

5.3.1 Direct Costs

The direct costs for 2500 t/d methanol plants utilizing the High Temperature Winkler process for lignite gasification are presented on Table 5-1 for the three processing scenarios.

5.3.2 Total Estimated Costs

The total estimated costs, comprising Total Direct Costs, Indirect Costs, Owner's Costs and Construction Camp Costs are shown on Table 5-2. The cost of installing a new 48 km long power transmission line to the Onakawana site from the existing provincial grid is included in the lignite/electrolytic hydrogen co-processing plant cost.

5.3.3 Operating Costs

The operating costs for the three processes are shown on Table 5-3 as follows:

- Lignite only with lignite at $15/tonne as received.
- Lignite/Natural Gas Co-processing with Natural Gas at $0.156/Nm^3. Included in the $58.3 mm operating cost is a demand charge of $9.2 mm to allow for construction of the new gas pipeline to the site approximately 180 km long and for maintenance of this pipeline.
- Lignite/Natural Gas Co-processing with Natural Gas at $.105/Nm^3. The same demand charge of $9.2 mm is applied to this cost.
- Lignite/electrolytic hydrogen co-processing with a power cost of 2.71c/kWh,
- Lignite/electrolytic hydrogen co-processing with a power cost of 1.8 c/kWh.

5.3.4 Summary of Capital and Operating Cost Estimates

This summary is shown on Table 5-4 and gives the following key costs per tonne of methanol based on an annual production of 800 000 tonnes.

- Total Estimated Capital Cost
- Total Operating Cost
- Total Production Cost
- Total Product Costs for capital recovery factors of 10, 15 and 20 percent.

5.4 Product Cost Sensitivity

Table 5-5 shows in formula manner, the sensitivity of product cost at 20 percent Capital Recovery Factor to costs of lignite, natural gas and electric power. For the lignite/natural gas co-processing option, the demand charge originally built into the operating cost is now reflected in the fixed cost of 194.38 shown.

TABLE 5-1

DIRECT COST - 2500 TONNE/DAY METHANOL PLANT

Estimated Costs in Million Canadian Dollars, 1st Quarter 1983.

SECTION	AREA	DESCRIPTION	LIGNITE ONLY	NAT. GAS CO. PROC.	ELECTROLYT H2
4100		LIGNITE PREPARATION			
	4110	Size Reduction	4.8	2.2	3.4
	4120	Drying	22.8	10.7	16.3
4200		GASIFICATION			
	4210	Gasifiers			
	4220	Waste Heat Boilers	161.6	81.3	118.4
	4230	Gas Quench			
4300		GAS COMPRESSION & TREATING			
	4310	Raw Gas Compression	20.9	8.7	15.7
	4320	CO Shift Conversion	18.3	----	5.6
	4330	Acid Gas Removal	60.0	21.9	29.9
4400		METHANOL SYNTHESIS & PURIFICATION			
	4410	Methanol Synthesis	71.0	71.0	71.0
	4420	Methanol Purification	1.0	1.0	1.0
4500		AIR SEPARATION, GAS CONVERSION, ELECTROLYSIS			
	4510	Air Separation	59.1	24.0	----
	4520	Steam Methane Reforming	----	59.5	----
	4530	Partial Oxidation	----	----	----
	4540	Pressure Swing Adsorption	----	----	----
	4550	Electrolysis Plant	----	----	120.0*
4600		SULPHUR PROCESSING			
	4610	Sulphur Recovery	6.3	3.0	4.4
	4620	Sulphur Slating	1.8	0.9	1.3
	4630	Tail Gas Processing	4.3	2.0	3.0
4700		GAS LIQUOR SEPARATION	----	----	----
		SUB TOTAL PROCESS UNITS	431.9	286.2	390.0
4800		EFFLUENT TREATMENT & DISPOSAL			
	4810	Fines Settling, Ash Dewatering	Incl in 4200	Incl in 4200	Incl in 4200
	4820	Solids Disposal	5.2	2.8	3.8
	4830	Ammonia Recovery	5.6	3.0	4.2
	4840	Wastewater Treatment	10.9	7.3	8.4
4900		UTILITIES, BUILDINGS & OFFSITES			
	4910	Steam Generation	28.9	17.0	22.9
	4920	Power Distribution	6.2	4.8	5.6
	4930	Cooling Water System	10.6	6.2	7.0
	4940	Raw Water System	5.5	3.9	4.3
	4950	Boiler Feed Water System & Demineralized Water	10.0	5.2	14.9
	4960	Firewater	2.0	2.0	2.0
	4970	Buildings	8.6	5.7	7.8
	4980	Offsites	34.5	34.5	34.5
		SUB TOTAL UTILITIES & SERVICES	128.0	92.4	115.4
		TOTAL	559.9	378.6	505.4
		SITE DEVELOPMENT @ 5%	28.0	18.9	25.3
		TOTAL DIRECT COST	587.9	397.5	530.7

* Includes H_2 and O_2 compression.

TABLE 5-2
2500 TONNE/DAY METHANOL PLANT
SUMMARY OF TOTAL ESTIMATED COSTS
(Million Canadian Dollars)

UNIT NO.	DESCRIPTION	LIGNITE ONLY	LIGNITE/ NAT. GAS CO. PROC.	LIGNITE/ ELEC. H2 CO. PROC.
100	Lignite Preparation	27.6	12.9	19.7
200	Gasification	161.6	81.3	118.4
300	Gas Compression & Treating	99.2	30.6	51.2
400	Methanol Synthesis & Purification	72.0	72.0	72.0
500	Air Separation, SMR, Electrolysis	59.1	83.5	120.0
600	Sulphur Processing	12.4	5.9	8.7
	SUB-TOTAL PROCESS UNITS	431.9	286.2	390.0
800	Effluent Treatment & Disposal	21.7	13.1	16.4
900	Utilities, Buildings & Offsites	106.3	79.3	99.0
	SUB-TOTAL UTILITIES & SERVICES	128.0	92.4	115.4
	TOTAL	559.9	378.6	505.4
	Site Development @ 5%	28.0	18.9	25.3
	TOTAL DIRECT COST	587.9	397.5	530.7
	INDIRECT COSTS			
	Engineering & Construction Fees @ 20%	117.6	79.5	106.1
	Contingency @ 14%	82.3	55.7	74.3
	TOTAL INSTALLED COST	787.8	532.7	711.1
	Power Line for Lignite/Electrolytic H2 Co-Pro.	----	----	10.8
	Owners Costs, Catalyst & Chemicals	15.8	10.6	14.2
	Construction Camp	55.9	37.8	50.4
	TOTAL ESTIMATED COST	859.5	581.1	786.5

TABLE 5-3
2500 TONNE/DAY METHANOL PLANT
ANNUAL OPERATING COSTS
(Million Canadian Dollars)

Item	Lignite Only	Lignite/ Nat. Gas Co-Processing Gas Cost \$0.156/Nm^3	Lignite/ Nat. Gas Co-Processing Gas Cost \$0.105/Nm^3	Lignite/ Electrolytic H2 Co-Processing Power Cost 2.71¢/KwH	Lignite/ Electrolytic H2 Co-Processing Power Cost 1.8¢/KwH
Operating Factor (Days)	320	320	320	320	320
Raw Materials					
- Lignite	37.2	12.6	12.6	23.0	23.0
- Natural Gas	----	58.3	42.2	----	----
Labour					
- Operating Labour	3.4	3.4	3.4	3.4	3.4
- Plant Overhead	1.0	1.0	1.0	1.0	1.0
- Local Admin.	.7	.7	.7	.7	.7
- Corp. Overhead	.3	.3	.3	.3	.3
- Burden and Supplies	1.4	1.4	1.4	1.4	1.4
Insurance	3.9	2.7	2.7	3.5	3.5
Taxes	2.0	1.4	1.4	1.8	1.8
Maintenance	13.0	10.4	10.4	10.6	10.6
Catalyst & Chemicals	8.6	5.7	5.7	5.4	5.4
Utilities					
- Power	3.6	3.0	3.0	87.8	58.3
- Lignite	10.5	5.4	5.4	5.7	5.7
- Natural Gas	----	11.0	7.8	----	----
TOTAL	85.6	117.3	98.0	144.6	115.1

TABLE 5-4

2500 TONNE/DAY METHANOL PLANT

SUMMARY OF CAPITAL AND OPERATING COST ESTIMATES

CASE	LIGNITE ONLY	LIGNITE/ NATURAL GAS $\$0.156/\text{Nm}^3$	LIGNITE/ NATURAL GAS $\$.105/\text{Nm}^3$	LIGNITE/ ELECTROLYTIC H2 2.71¢/KwH	LIGNITE/ ELECTROLYTI H2 1.8¢/KwH
CAPITAL COST $MM CAN.					
Process Units	431.9	286.2	286.2	390.0	390.0
Other Services and Site Dev.	156.0	111.3	111.3	140.7	140.7
Engineering, Indirects, Contingency	199.9	135.2	135.2	180.4	180.4
TOTAL INSTALLED COST	787.8	532.7	532.7	711.1	711.1
Owners Costs, Catalysts, Camp	71.7	48.4	48.4	64.6	64.6
Power Line for Electrolytic Option				10.8	10.8
TOTAL ESTIMATED COST $MM	859.5	581.1	581.1	786.5	786.5
TOTAL ESTIMATED COST $/t METHANOL	1074.4	726.4	726.4	983.1	983.1
OPERATING COSTS $/t					
Raw Materials - Lignite	46.50	15.75	15.75	28.75	28.75
- Natural Gas	----	72.87	52.75	----	----
Labour, Overhead, Admin., Supplies	8.50	8.50	8.50	8.50	8.50
Insurance and Municipal Taxes	7.37	5.13	5.13	6.62	6.62
Maintenance	16.25	13.00	13.00	13.25	13.25
Catalysts and Chemcials	10.75	7.13	7.13	6.75	6.75
Utilities - Power	4.50	3.75	3.75	109.75	72.87
- Lignite	13.12	6.75	6.75	7.12	7.12
- Natural Gas	----	13.75	9.75	----	----
TOTAL OPERATING COST $/TONNE	106.99	146.63	122.51	180.74	143.86
By Product Credits					
- Sulphur @ $100/t	2.40	0.80	0.80	1.48	1.48
- Ammonia @ $240/t	2.21	0.77	0.77	1.34	1.34
TOTAL BY-PRODUCTS CREDIT	4.61	1.57	1.57	2.82	2.82
TOTAL PRODUCTION COST $/TONNE	102.38	145.06	120.94	177.92	141.04
(Operating Cost minus By-Product Credit)					
Total Product Cost, $/Tonne (Production Cost + (Total Est. Cost/t x percentage Capital Recovery)					
at 10% Capital Recovery	209.82	217.70	193.58	276.23	239.35
at 15% Capital Recovery	263.54	254.02	229.90	325.38	288.50
at 20% Capital Recovery	317.26	290.34	266.22	374.54	337.66

TABLE 5-5

2500 TONNE/DAY METHANOL PLANT

TOTAL PRODUCT COST SENSITIVITY TO RAW MATERIALS & UTILITIES COSTS

Basis: 20 percent Capital Recovery Factor
Product Cost expressed in \$/tonne Methanol
Lignite Cost (L) expressed in \$/tonne (as received basis)
Natural Gas Cost (NG) expressed in \$/Nm^3
Power Cost (E) expressed in cents/KwHr

Lignite Only

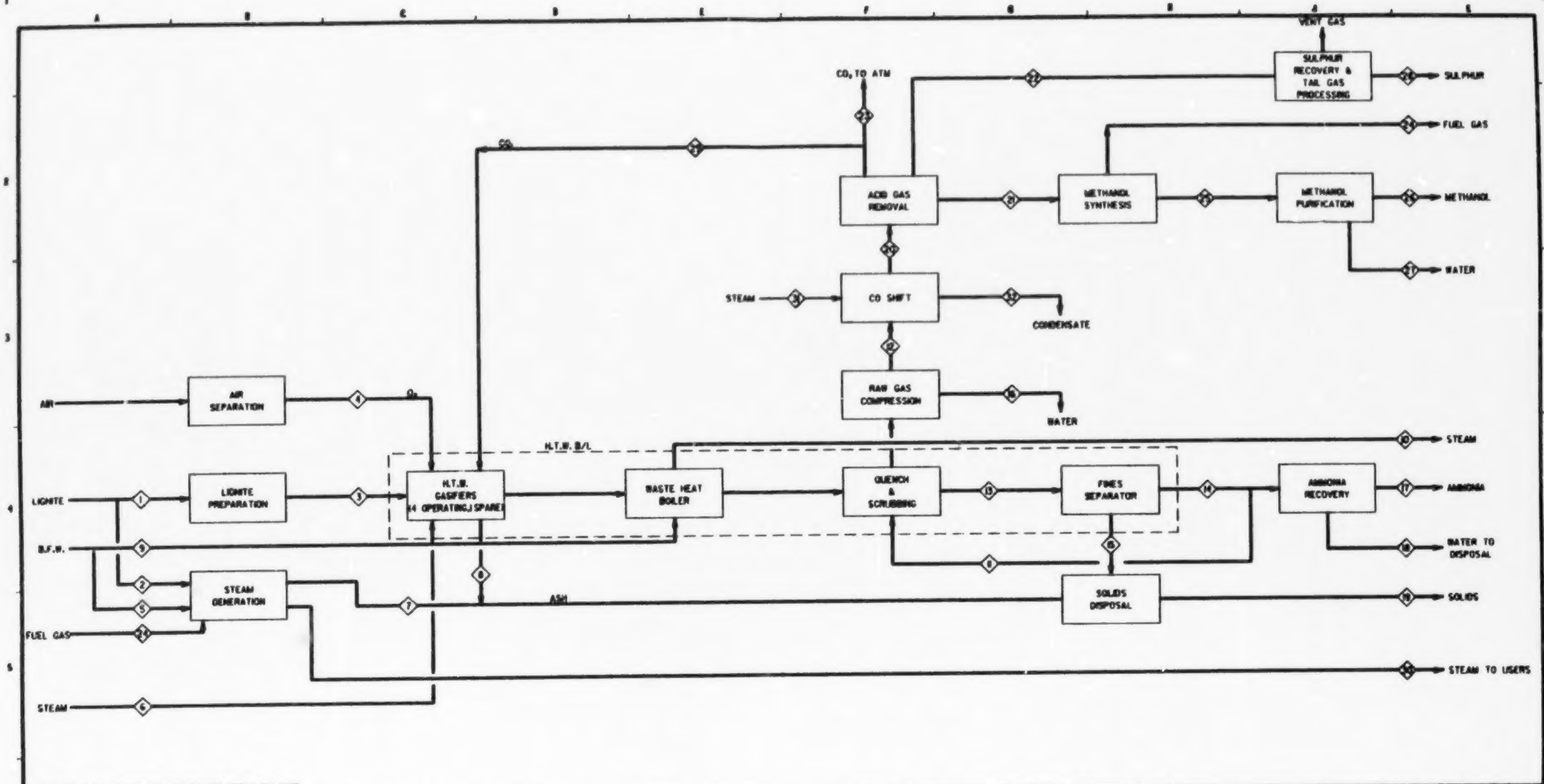
Product Cost: 257.64 + 3.975 L

Lignite/Natural Gas Co-Processing

Product Cost: 194.38 + 1.5 L + 470.8 NG

Lignite/Electrolytic Hydrogen Co-Processing

Product Cost: 228.92 + 2.39 L + 40.5 E


AIR
AIR SEPARATION
O_2
LIGNITE
LIGNITE PREPARATION
B.F.W.
STEAM GENERATION
FUEL GAS
STEAM
H.T.W. GASIFIERS (4 OPERATING,1 SPARE)
H.T.W. B/L
ASH
CO_2
WASTE HEAT BOILER
QUENCH & SCRUBBING
FINES SEPARATOR
SOLIDS DISPOSAL
AMMONIA RECOVERY
RAW GAS COMPRESSION
WATER
CO SHIFT
CONDENSATE
ACID GAS REMOVAL
CO_2 TO ATM
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
VENT GAS
SULPHUR
FUEL GAS
METHANOL
WATER
STEAM
AMMONIA
WATER TO DISPOSAL
SOLIDS
STEAM TO USERS
1
2
3
4
5
6
7
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11
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NOTE: REFER TO DWG.NO.400-43-008 FOR STEAM BALANCE.

STREAM NO.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP. (MAX MOIST.)	AS RECEIVED LIGNITE TO STEAM PLANT. (MAX MOIST.)	LIGNITE TO GASIFIER (12% MOIST.)	OXYGEN TO GASIFIER	B.F.W. TO STEAM PLANT	STEAM TO GASIFIER	ASH FROM STEAM PLANT TO DISPOSAL	ASH FROM GASIFIER TO DISPOSAL	B.F.W.	M.P. STEAM (3.5 MPa)	QUENCH WATER	RAW GAS	QUENCH WATER WITH FINES	SEPARATED WATER TO NH_3 RECOVERY	SLURRY TO SOLIDS DISPOSAL	WATER FROM COMPRESS-ION
FLOW RATE t/d	7748	2805	4842	2533	10544	1038	225	761	4037	3958	888	7407	8995	1504	39	27

STREAM NO.	17	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32
DESCRIPTION	ANHYDROUS AMMONIA	WATER TO DISPOSAL	SOLIDS TO LANDFILL	SHIFTED SYNGAS	SYNGAS TO METHANOL SYNTHESIS	ACID GAS TO CLAUS PLANT	CO_2 TO ATM.	FUEL GAS	RAW METHANOL TO PURIFICATION	FUEL GRADE METHANOL	WATER	SULPHUR	CO_2 TO GASIFIER	HP STEAM TO USERS (10.0 MPa)	MP STEAM TO SHIFT (3.5 MPa)	CONDENSATE FROM SHIFT
FLOW RATE t/d	23	653	1317	8590	2933	64	4743	360	2572	2500	72	60	850	10337	2383	1200

C	RELEASED FOR PHASE 2A FINAL REPORT		CM	
B	RELEASED FOR PHASE 2A DRAFT		CM	
A	RELEASED FOR PHASE 2 REPORT		CM	
	DESCRIPTION			

REVISIONS

REVISIONS

REFERENCE DRAWINGS

ONTARIO ENERGY CORPORATION

TORONTO, ONTARIO

KILBORN

HIGH TEMPERATURE WINKLER

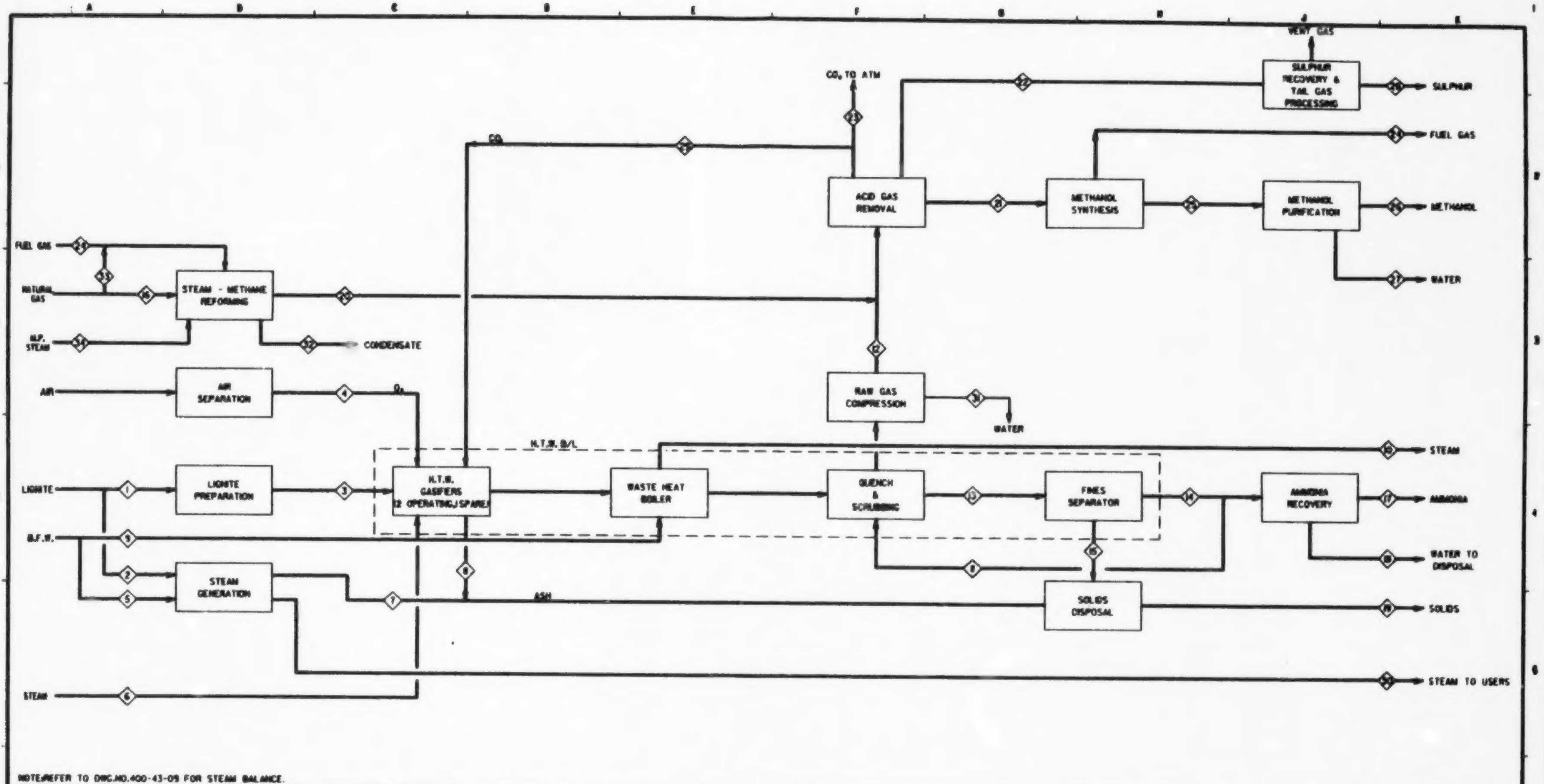
BASE CASE

LIGNITE ONLY

3484

400-43-05

C



FUEL GAS
NATURAL GAS
H.P. STEAM
AIR
LIGNITE
B.F.W.
STEAM
STEAM - METHANE REFORMING
CONDENSATE
AIR SEPARATION
O₂
LIGNITE PREPARATION
STEAM GENERATION
H.T.W. GASIFIERS (2 OPERATING,1 SPARE)
H.T.W. B/L
CO₂
WASTE HEAT BOILER
ASH
CO₂ TO ATM
ACID GAS REMOVAL
RAW GAS COMPRESSION
WATER
QUENCH & SCRUBBING
FINES SEPARATOR
SOLIDS DISPOSAL
METHANOL SYNTHESIS
METHANOL PURIFICATION
VENT GAS
SULPHUR RECOVERY & TAIL GAS PROCESSING
AMMONIA RECOVERY
SULPHUR
FUEL GAS
METHANOL
WATER
STEAM
AMMONIA
WATER TO DISPOSAL
SOLIDS
STEAM TO USERS
1
2
3
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5
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34

NOTE:REFER TO DWG.NO.400-43-09 FOR STEAM BALANCE.

STREAM NO.	DESCRIPTION	FLOW RATE (t/d)
1	AS RECEIVED LIGNITE TO LIGNITE PREP. (45% MOIST.)	2630
2	AS RECEIVED LIGNITE TO STEAM PLANT. (45% MOIST.)	820
3	LIGNITE TO GASIFIER (22% MOIST.)	1644
4	OXYGEN TO GASIFIER	860
5	B.F.W. TO STEAM PLANT	3650
6	STEAM TO GASIFIER	352
7	ASH FROM STEAM PLANT TO DISPOSAL	116
8	ASH FROM GASIFIER TO DISPOSAL	258
9	B.F.W.	1371
10	H.P. STEAM (3.5 MPa)	1344
11	QUENCH WATER	281
12	RAW GAS	2510
13	QUENCH WATER WITH FINES	644
14	SEPARATED WATER TO NH₃ RECOVERY	58
15	SLUDGE TO SOLIDS DISPOSAL	133
16	NATURAL GAS TO REFORMER	727
17	ANHYDROUS AMMONIA	8
18	WATER TO DISPOSAL	222
19	SOLIDS TO LANDFILL	507
20	REFORMED GAS	1696
21	SYNGAS TO METHANOL SYNTHESIS	2866
22	ACID GAS TO CLAUS PLANT	22
23	CO₂ TO ATM.	1030
24	FUEL GAS	294
25	RAW METHANOL TO PURIFICATION	2572
26	FUEL GRADE METHANOL	2500
27	WATER	72
28	SULPHUR	20
29	CO₂ TO GASIFIER	288
30	HP STEAM TO USERS (10.0 MPa)	3570
31	WATER FROM COMPRESS-ION	14
32	CONDEN-SATE	886
33	NATURAL GAS TO FUEL GAS	143
34	PROCESS STM TO REFORMER	2136

REFERENCE DRAWINGS

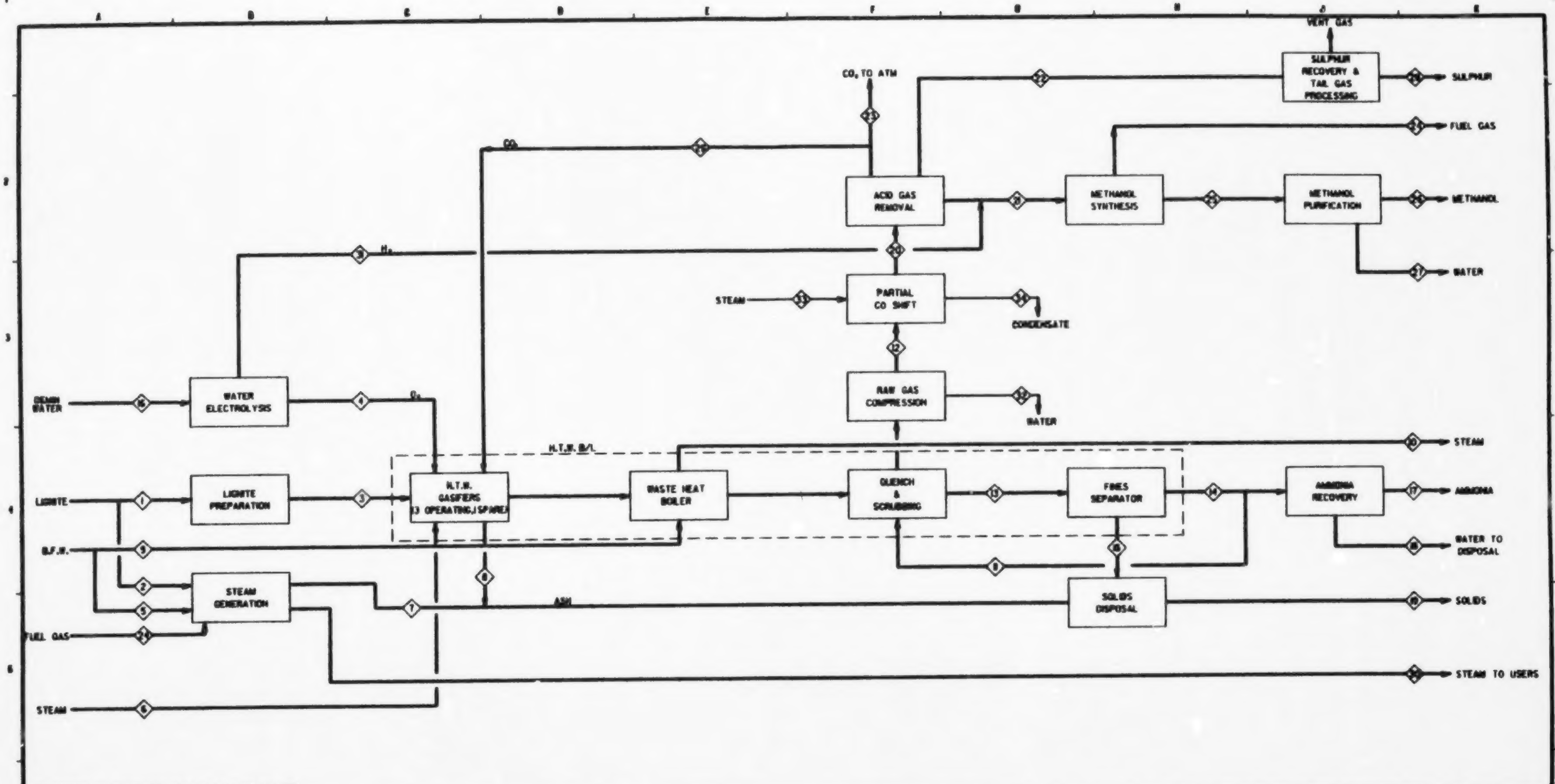
REV.	DESCRIPTION		
C	RELEASED FOR PHASE 2A FINAL REPORT	[illegible]	GM
B	RELEASED FOR PHASE 2A DRAFT	[illegible]	GM
A	RELEASED FOR PHASE 2 REPORT	2/83	GM

REVISIONS

ONTARIO ENERGY CORPORATION
TORONTO , ONTARIO
KILBORN

HIGH TEMPERATURE WINKLER
LIGNITE / NATURAL GAS
CO - PROCESSING

3494
17
400-43-06
C


DEMIN WATER
LIGNITE
B.F.W.
FUEL GAS
STEAM
WATER ELECTROLYSIS
LIGNITE PREPARATION
STEAM GENERATION
O2
H2
CO2
H.T.W. GASIFIERS (3 OPERATING, 1 SPARE)
H.T.W. B/L
WASTE HEAT BOILER
ASH
QUENCH & SCRUBBING
FINES SEPARATOR
SOLIDS DISPOSAL
AMMONIA RECOVERY
RAW GAS COMPRESSION
WATER
PARTIAL CO SHIFT
CONDENSATE
ACID GAS REMOVAL
CO2 TO ATM
METHANOL SYNTHESIS
METHANOL PURIFICATION
SULPHUR RECOVERY & TAIL GAS PROCESSING
VENT GAS
SULPHUR
FUEL GAS
METHANOL
WATER
STEAM
AMMONIA
WATER TO DISPOSAL
SOLIDS
STEAM TO USERS

NOTE: REFER TO DWG. NO. 400-43-10 FOR STEAM BALANCE.

STREAM NO.	1	2	3	4	5	6	7	8	9	10	11	12	13	14	15	16	17
DESCRIPTION	AS RECEIVED LIGNITE TO LIGNITE PREP.	AS RECEIVED LIGNITE TO STEAM PLANT.	LIGNITE TO GASIFIER	OXYGEN TO GASIFIER	B.F.W. TO STEAM PLANT	STEAM TO GASIFIER	ASH FROM STEAM PLANT TO DISPOSAL	ASH FROM GASIFIER TO DISPOSAL	B.F.W.	H.P. STEAM (3.5 MPa)	QUENCH WATER	RAW GAS	QUENCH WATER WITH FINES	SEPARATED WATER TO NH$_3$ RECOVERY	SLUDGE TO SOLIDS DISPOSAL	FEED WATER TO WATER ELECTROLYSIS	ANHYDROUS AMMONIA
[illegible]	4792	894	2995	1559	6656	642	123	471	2497	2449	505	4574	865	923	242	1754	14

STREAM NO.	18	19	20	21	22	23	24	25	26	27	28	29	30	31	32	33	34
DESCRIPTION	WATER TO DISPOSAL	SOLIDS TO LANDFILL	SHIFTED SYNGAS	SYNGAS TO METHANOL SYNTHESIS	ACID GAS TO CLAUS PLANT	CO$_2$ TO ATM.	FUEL GAS	RAW METHANOL TO PURIFICATION	FUEL GRADE METHANOL	WATER	SULPHUR	CO$_2$ TO GASIFIER	HP STEAM TO USERS	HYDROGEN FROM ELECTROLYSIS	WATER FROM COMPRESSION	HP STEAM TO SHIFT (3.5 MPa)	CONDENSATE FROM SHIFT
[illegible]	404	835	4753	2879	39	1804	307	2572	2500	72	37	525	6825	185	17	369	190

C	RELEASED FOR PHASE 2A FINAL REPORT		CM
B	RELEASED FOR PHASE 2A DRAFT		CM
A	RELEASED FOR PHASE 2 REPORT		CM
	DESCRIPTION		

REFERENCE DRAWINGS

REVISIONS

ONTARIO ENERGY CORPORATION

TORONTO, ONTARIO

KILBORN

HIGH TEMPERATURE WINKLER

LIGNITE / ELECTROLYTIC H$_2$

CO - PROCESS

3484

400-43-07 C

APPENDIX 2

RESULTS OF THE GASIFICATION TESTS CARRIED OUT ON ONTARIO AND SASKATCHEWAN LIGNITES

By

RHEINBRAUN-Consulting, GmbH

GASIFICATION IN THE
HTW PROCESS DEVELOPMENT UNIT

Report on the results of gasification tests carried out
in the HTW PDU with Ontario Lignite and Saskatchewan Lignite
for the Company Ontario Energy Corporation

by

Andreas Bellin
Lothar Schrader

Cologne, October 1983
Department: VF 2/3

NOTE: The original report prepared by RHEINBRAUN contains confidential information which does not appear in this issue.

Introduction

On the basis of an agreement concluded between the company Ontario Energy Corporation and Rheinbraun-Consulting tests on the gasifiability of Ontario-Lignite and Saskatchewan-Lignite were made. The test program to be performed by Rheinbraun-Consulting dealt with the following:

1. Physicochemical Tests

   The main physicochemical tests on the lignites were carried out by OEC according to a.m. agreement. The results are shown in Table 1. So Rheinbraun performed the following tests:

   - reactivity
   - pore volume
   - ignition temperature
   - screen analysis

2. One fluidized bed test at normal pressure and ambient temperature for each of the crushed and dried lignites.

3. Gasification Tests

   One gasification test applying oxygen/steam as gasifying agents for each of the two lignites.

1 Physicochemical Tests

1.1 Reactivity

The measurements of reactivity were made with coked lignite in the equipment shown in Figure 1.

In the reactivity measurements, heat was added by having N_2 passed through the equipment at 5°C/min till the test temperature was reached; subsequently, CO_2 passed through the sample at constant temperatures. The exhaust gas leaving the reaction tube was continuously analyzed for its CO_2 content in a CO_2-analyzer.(1)

The reactivity measurements were carried out under the following conditions:

quantity of gas	9 L CO_2 per hour
test portion	3 g C_{fix}
temperatures	850, 900, 950°C

The results of the reactivity measurements are shown in the following figures:

Fig.2: C-conversion over the test period at the selected temperature stages for Ontario-Lignite

Fig.3: C-conversions over the test period at the selected temperature stages for Saskatchewan-Lignite

Fig.4: ln k_m over 1/T (Arrhenius diagram) for k_m at 30% C-conversion for both lignites

The velocity constant is a function of temperature according to the Arrhenius equation

$$k_m = k_o \exp(-E_A/RT)$$

where k_m is the velocity constant,
k_o the frequency factor,
E_A the activation energy,
R the general gas constant,
T the temperature.

The following table contains the data on reaction kinetics:

Temp. (°C)	Velocity constant k_m (cm^3/g-s) (at a C-conversion of 30%)	
	Ontario-Lignite	Saskatchewan-Lignite
850	1.46	1.88
900	3.29	4.42
950	7.69	11.83

	Ontario-Lig.	Sask-Lig.
Activity energy E_A (kJ/mol)	182.0	199.5
Frequency factor k_o (cm^3/g-s)	4.35×10^8	3.69×10^9

A comparison of the k_m values at the various temperature stages reveals an increase in k_m by a factor of some 2.2-2.7 with an increase in temperature of 50°C. In their levels, the E_A and k_o values are similar to other carbon carriers, such as North American lignite and Rhenish brown coal studied so far.

1.2 Porosity Measurements

The distribution of pore volume for the lignites is shown in Figure 5.

The pore volume as a function of the pore radius was determined in the mercury porosimeter.

The sum of specific pore volume is

7.38×10^{-2} cm^3/g for Ontario-Lignite
4.98×10^{-2} cm^3/g for Saskatchewan-Lignite

1.3 Ignition Temperature

The ignition temperature was likewise determined in the equipment shown in Figure 1. In this process, the sample is heated up by 5°C/min passed through a stream of oxygen (9 L/h). The following ignition temperatures were determined:

Ontario-Lignite	coal	195°C
Ontario-Lignite	coke	310°C
Saskatchewan-Lignite	coal	195°C
Saskatchewan-Lignite	coke	285°C

1.4 Screen Analysis and Density Determination

The screen analysis and the density determination for the two crushed and dried lignites are shown in Table 2.

The average grain diameter is determined according to the following formula:

$$\overline{d} = \frac{1}{\sum (x/d)_i}$$

where x is the share of the grain fraction and d the average diamater of the grain fraction.

1.5 Comparison of Characteristical Data on Gasification of Ontario-Lignite and Saskatchewan-Lignite

Table 3 shows a comparison made between the two lignites.

2 Fluidization Tests in a Cold Model

Fluidization tests were made in a glass model having an inside diameter of 200 mm. Nitrogen served as fluidization medium at ambient temperature (20°C).

2.1 Ontario-Lignite

Figure 7 shows the grain size distribution as well as the dependence of the pressure drop and the fluidized bed height on the fluidization velocity. A gas velocity of 15 cm/s results for the minimum fluidization point. Visual observations, however, proved that a sufficiently homogenous fluidization takes place only at gas velocities of more than 20 cm/s.

The quantity of the sample tested amounted to 10.0 kg.

2.2 Saskatchewan-Lignite

Figure 8 shows the grain size distribution as well as the dependence of the pressure drop and the fluidized bed weight on the fluidization velocity. A minimum fluidization velocity of 32 cm/s results. A sufficiently homogenous fluidization takes place at gas velocities of more than 45 cm/s.

The quantity of the sample tested amounted to 14.3 kg.

The different minimum fluidization velocities of the lignites are a result of the diverse crushing behaviour in the hammer-mill.

3 Gasification Tests

3.1 General Remarks

The gasification tests were carried out with the lignites in the plant shown in Figure 9. Figure 10 shows the flow sheet of the plant.

The gasification tests were carried out with oxygen and steam as main gasification agent.

Start-up of the plant was carried out according to the following method:

Prior to the test, the reactor was heated by preheated air for approximately four hours.

The temperature was initially increased with air to about 650°C. Then the operation was switched to oxygen/steam. The oxygen/steam ratio was changed stepwise to the required test conditions. For an exact determination of the product gas quantity, a minimum throughput of air had to be maintained during the test.

The reactor temperatures were registered by a multi-channel recorder.

3.2 Gasification Tests with Ontario-Lignite

The test duration was limited to two hours after obtaining the given temperatures. During these two hours, gas samples were taken and analyzed every 20 minutes. Figure 12 shows the course of temperature, pressure drop through the fluidized bed and product gas composition.

The ash was discharged continuously by a screw in an airtight container. The high amount of dust (3.4 kg/h) discharged by the second cyclone is mainly a result of the small grain size of the crushed coal. For further tests, the lignite should be crushed to a more coarse grain size. Within the two hours of steady state run, no problems with ash agglomerates occurred. After these two hours, the oxygen injected in the entrained phase was increased from 4 to

5 m^3 (STP)/h, in order to test the influence on the ash fusion. The average temperature in the entrained phase rose from 895°C to 975°C. The duration of this test was limited to 40 minutes, because the lignite was running short. When the reactor was opened, some minor ash agglomerates were found which, however, did not interfere with the gasifier operation.

3.3 Gasification Test with Saskatchewan-Lignite

The test duration was limited to two hours after obtaining the given temperatures. During these two hours, gas samples were taken and analyzed every 20 minutes. Figure 15 shows the course of temperature, pressure drop through the fluidized bed and product gas composition.

Because of better grain size of the lignite, only a small amount of dust (0.2 kg/h) was discharged by the second cyclone. But the amount of ash discharged by a screw was rather high (4.1 kg/h).

It was tried to test the coal under quite the same conditions as Ontario-Lignite. The average temperatures in the entrained bed were about 915°C and a maximum temperature of 990°C.

After opening the reactor, no ash agglomerates could be found.

4 Result Evaluation

4.1 Suitability of the Lignites

The two lignites have been demonstrated suitable for fluidized bed gasification despite their relatively low ash softening point and their rather high ash content.

4.2 Performance Data

The degree of C-conversion and the specific synthesis gas yield depending thereon are mainly determined both by the recirculation rate of the dust and the quantity of ash discharged from the fluidized bed.

The recirculation rate of the dust depends on the design of the first cyclone. Maximum adjustment of the cyclone to all specific gasifying conditions, however, would exceed the scope of such a test program.

In the process development unit, ash discharge from the fluidized bed meets the following problems: The ash is discharged continuously and fed by screw into an airtight container. Monitoring of the ash quantities discharged, however, is carried out discontinuously at intervals of about two hours. The sizes of the ash particulates require a certain minimum size of the discharge screw. Consequently, an adjustment of the quantity to be discharged can be effected only by screw speed. With these low speeds, an adjustment is difficult to obtain so that an optimized ash discharge is only feasible after further optimization steps in this process development unit.

4.3 CO_2 Content in the Raw Gas

Due to the unsatisfactory ratio of converted lignite to heat-radiating surface in the HTW process development unit, a comparably high heat loss to total energy flow ratio occurs (10% to 15% heat loss depending on the throughput). This fact has to be considered in evaluating the gas quality, in particular the high CO_2 content.

The relatively high moisture content of the raw gas is another cause of the high CO_2 content. Due to the portion of moisture already in the gasifier, a shift conversion reaction takes place according to the following equation:

$$CO + H_2O \rightleftharpoons CO_2 + H_2$$

This is proved by the high H_2/CO ratio in the raw gas.

References

1. K. Hedden, Über die Reaktionsfähigkeit von Koks, Brennstoff-Chemie 41 (1960), Nr. 7, S. 193-203.

2. Kunii and Levenspiel, Fluidization Engineering, Robert E. Krieger Publishing Co., Inc. 1977 S. 67-71.

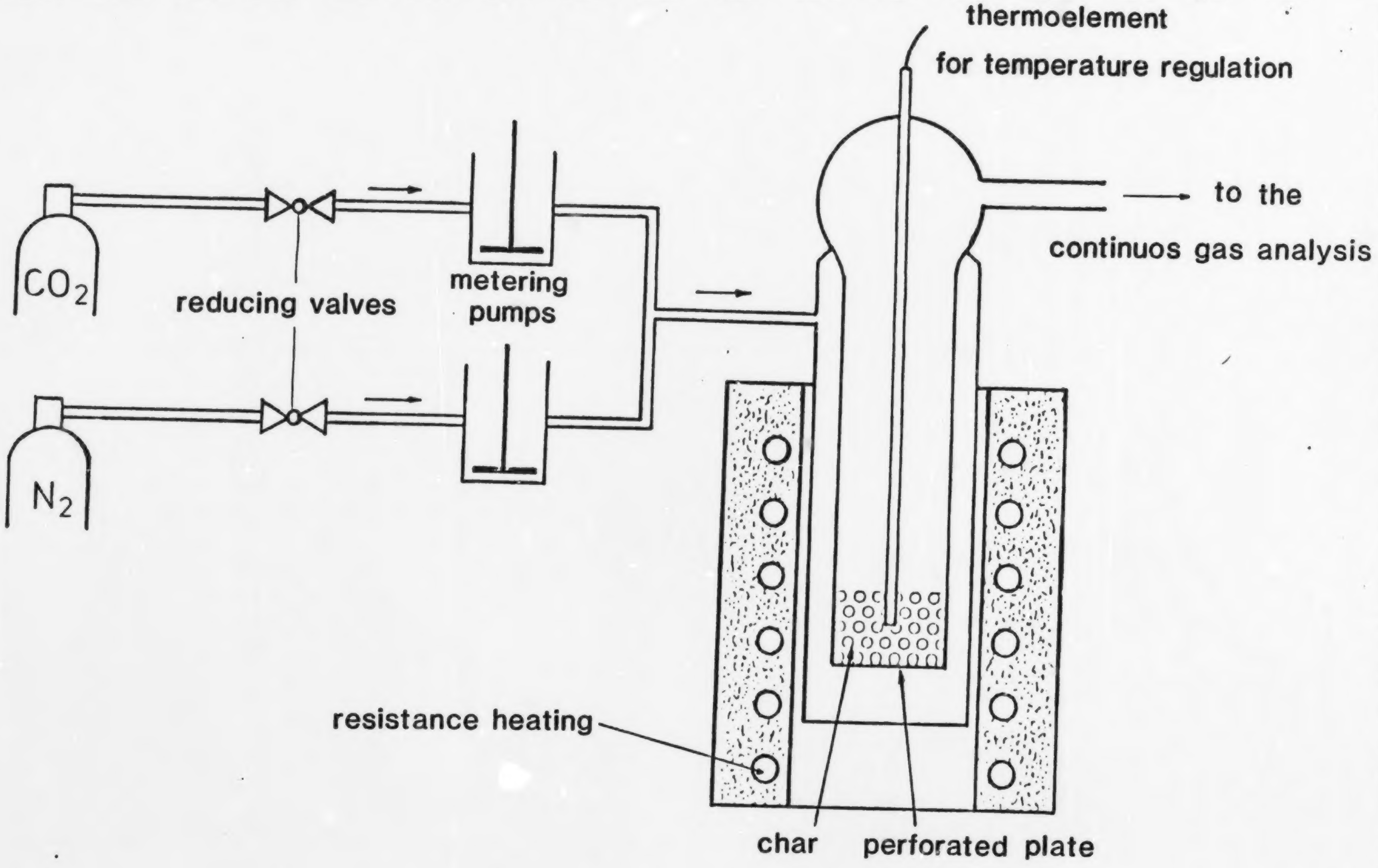

thermoelement
for temperature regulation
to the
continuos gas analysis
CO_2
N_2
reducing valves
metering
pumps
resistance heating
char
perforated plate

Fig. 1 : Equipment for determination of reaction velocity

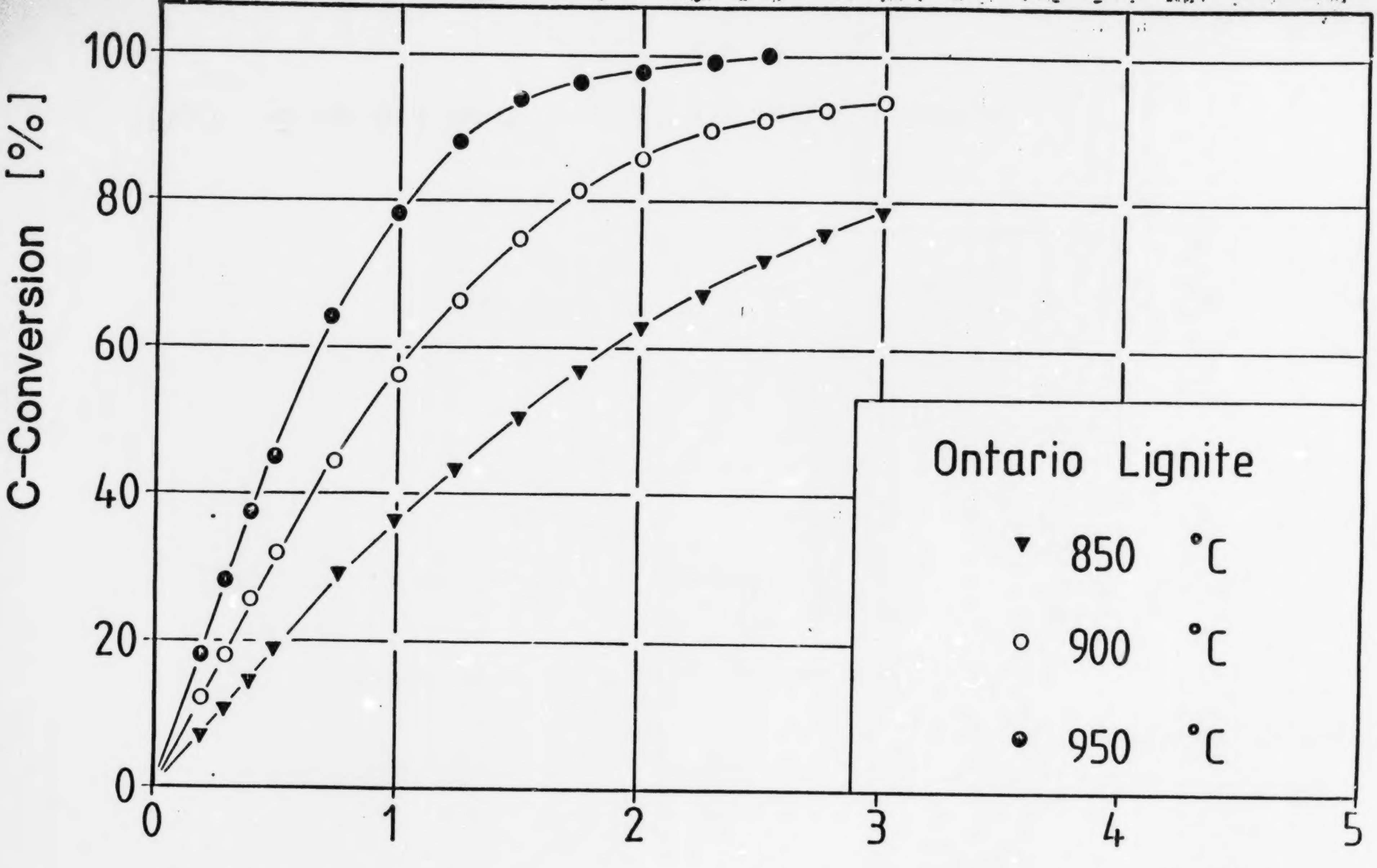


Fig.2

C-conversion over the test period at the selected temperature stages

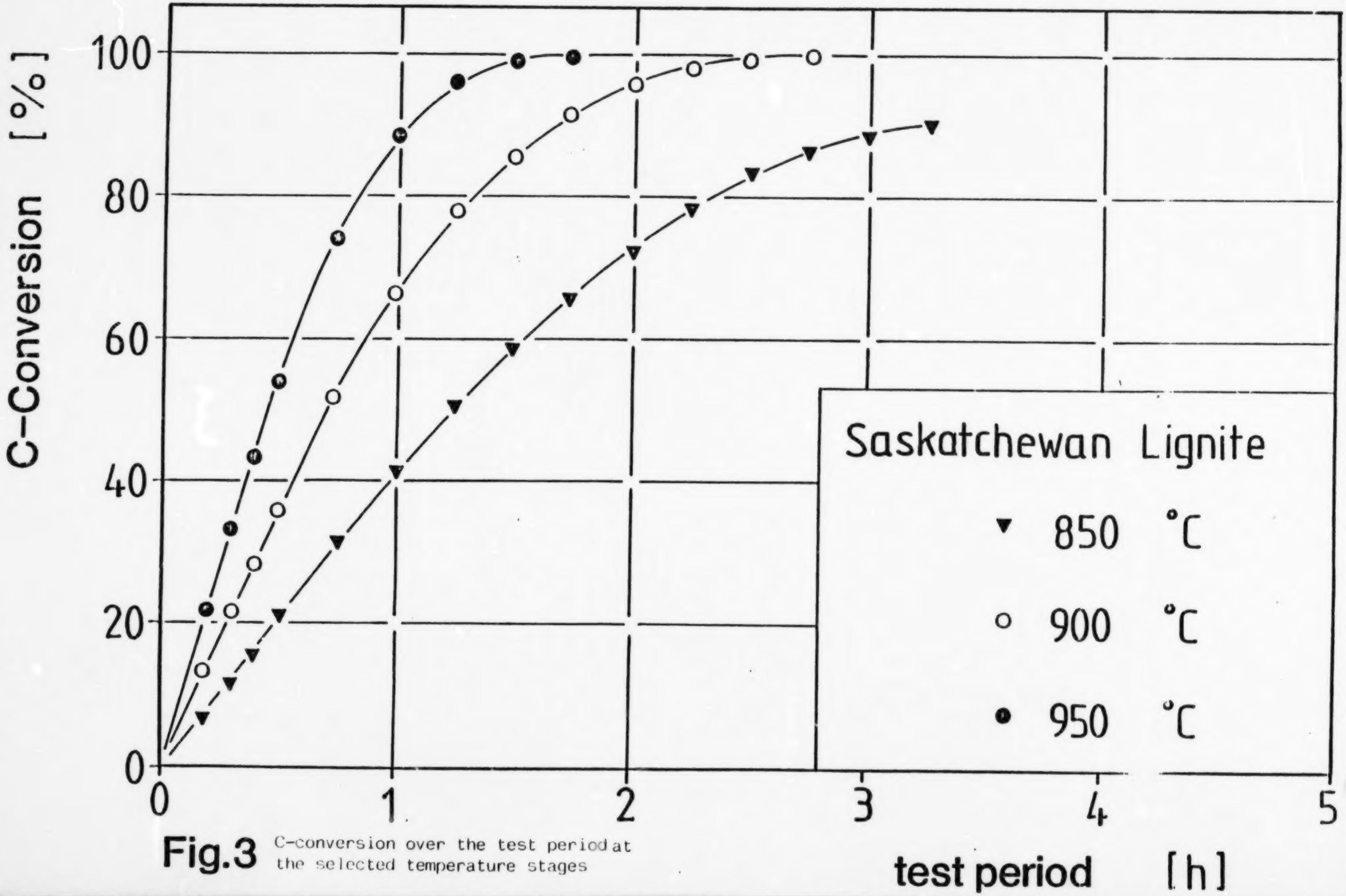


Fig.3 C-conversion over the test period at the selected temperature stages

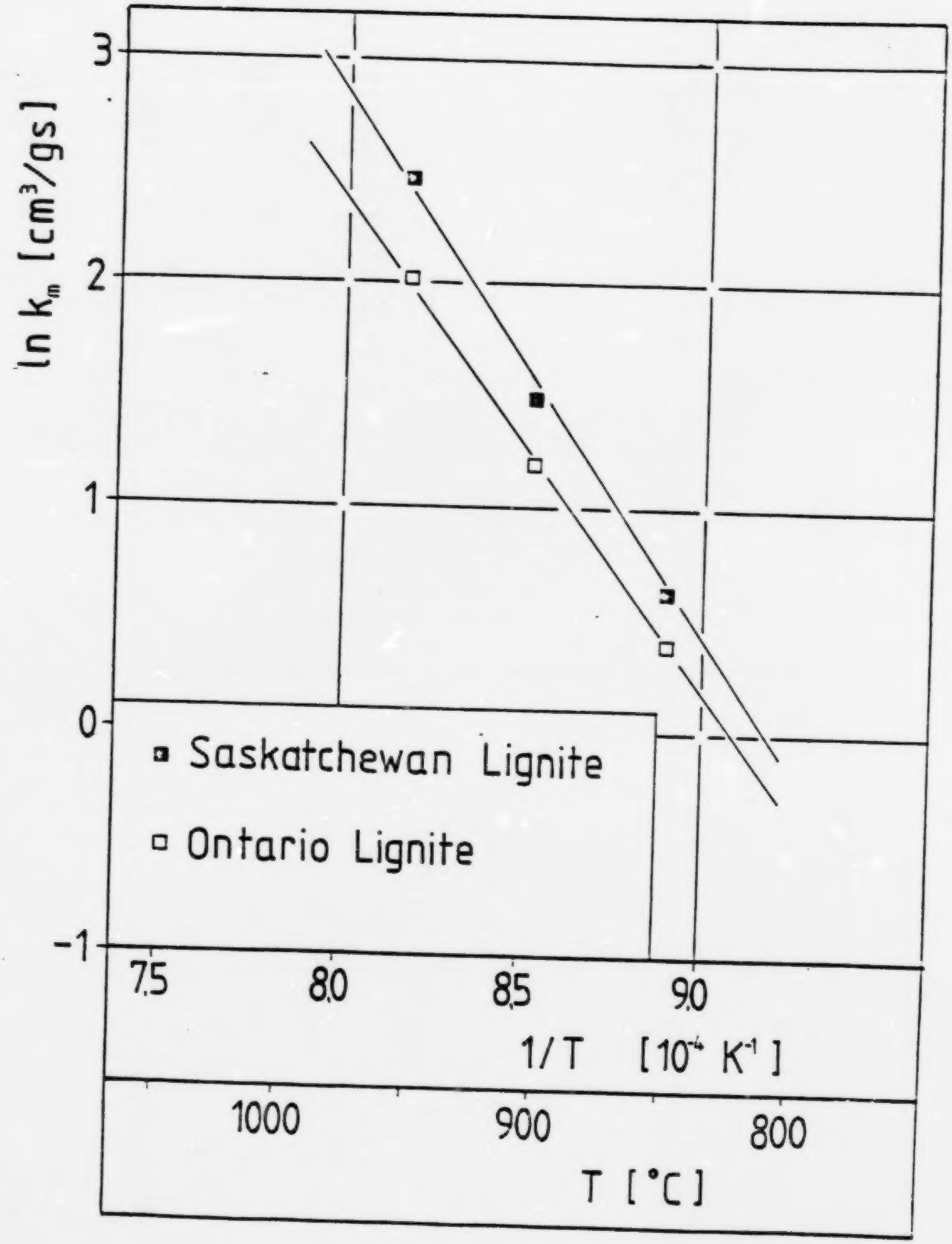


Fig.4 ln k_m versus 1/T (Arrhenius diagram)

k_m at 30 % C-conversion

Fig.5 distribution of pore volume

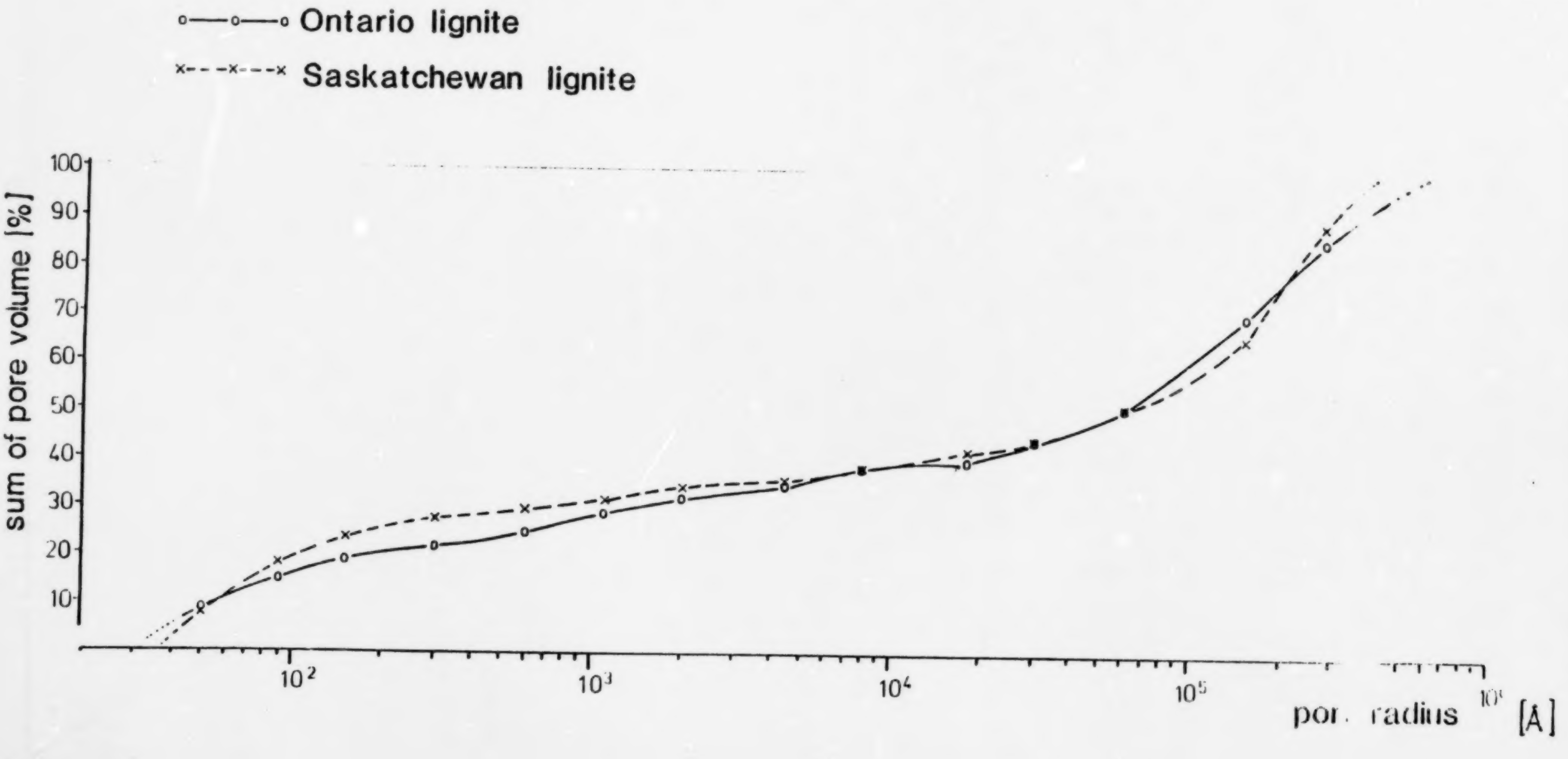
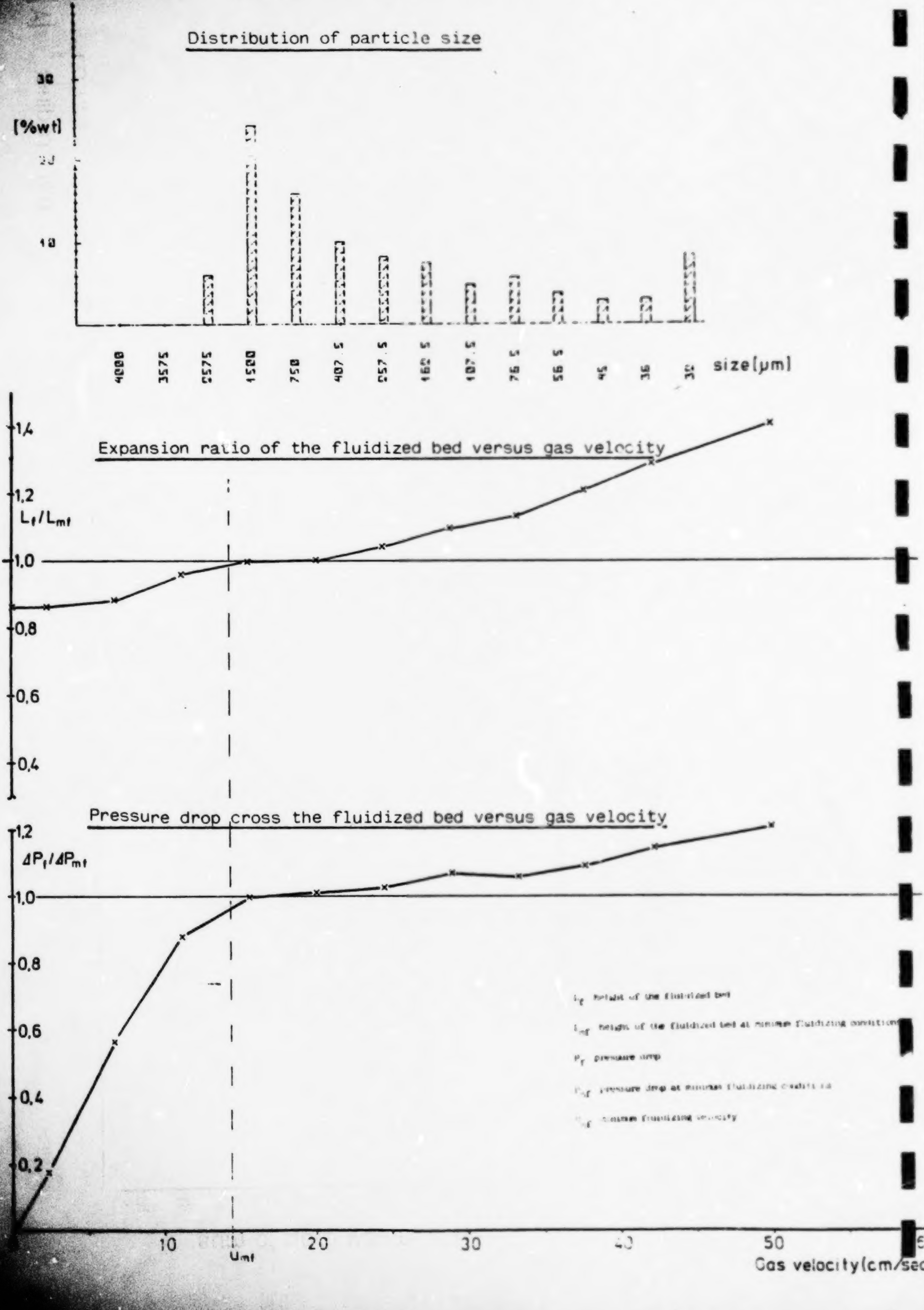


Fig.7 Ontario-lignite

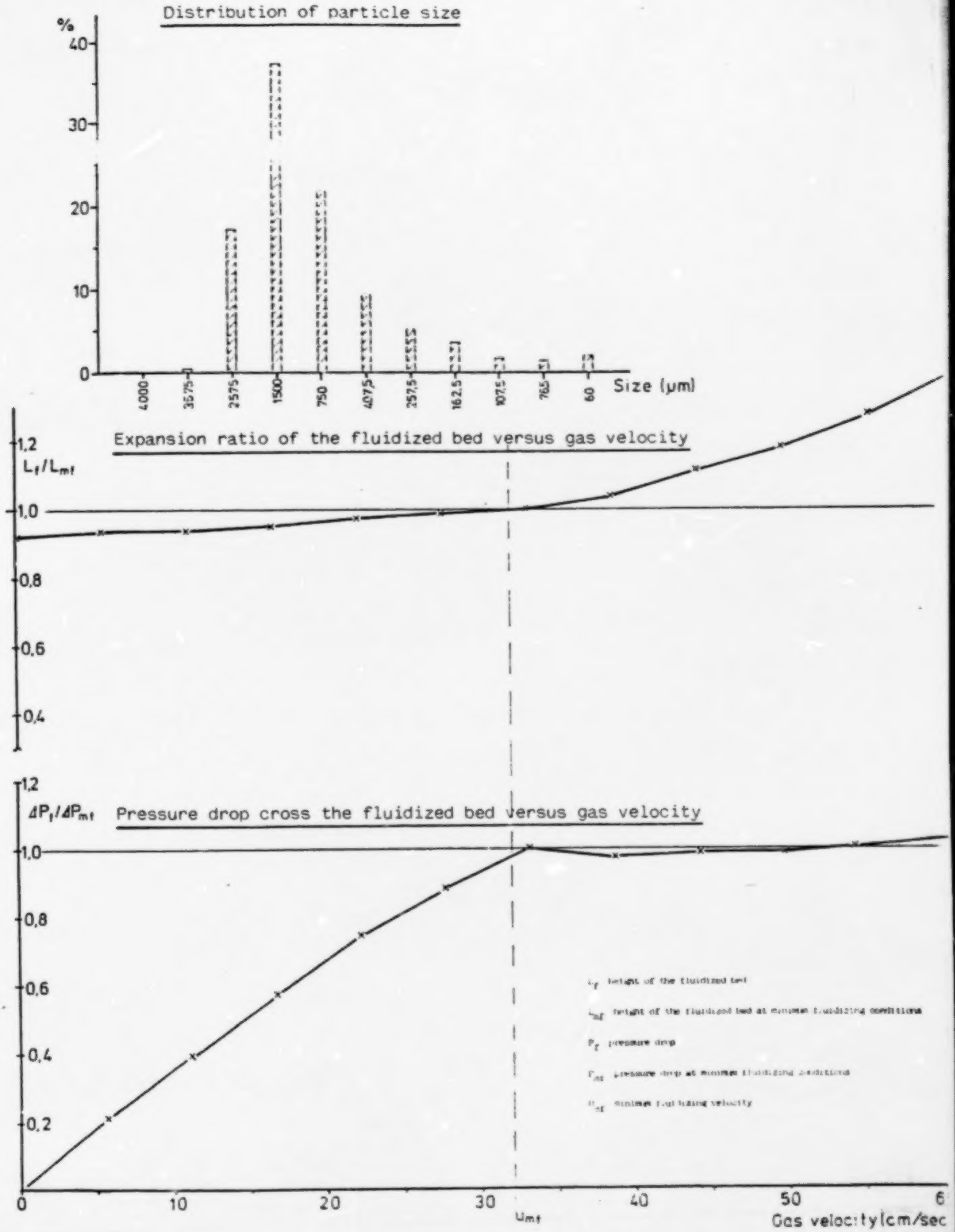


Distribution of particle size

Expansion ratio of the fluidized bed versus gas velocity

Pressure drop cross the fluidized bed versus gas velocity

Fig.8 Saskatchewan-lignite



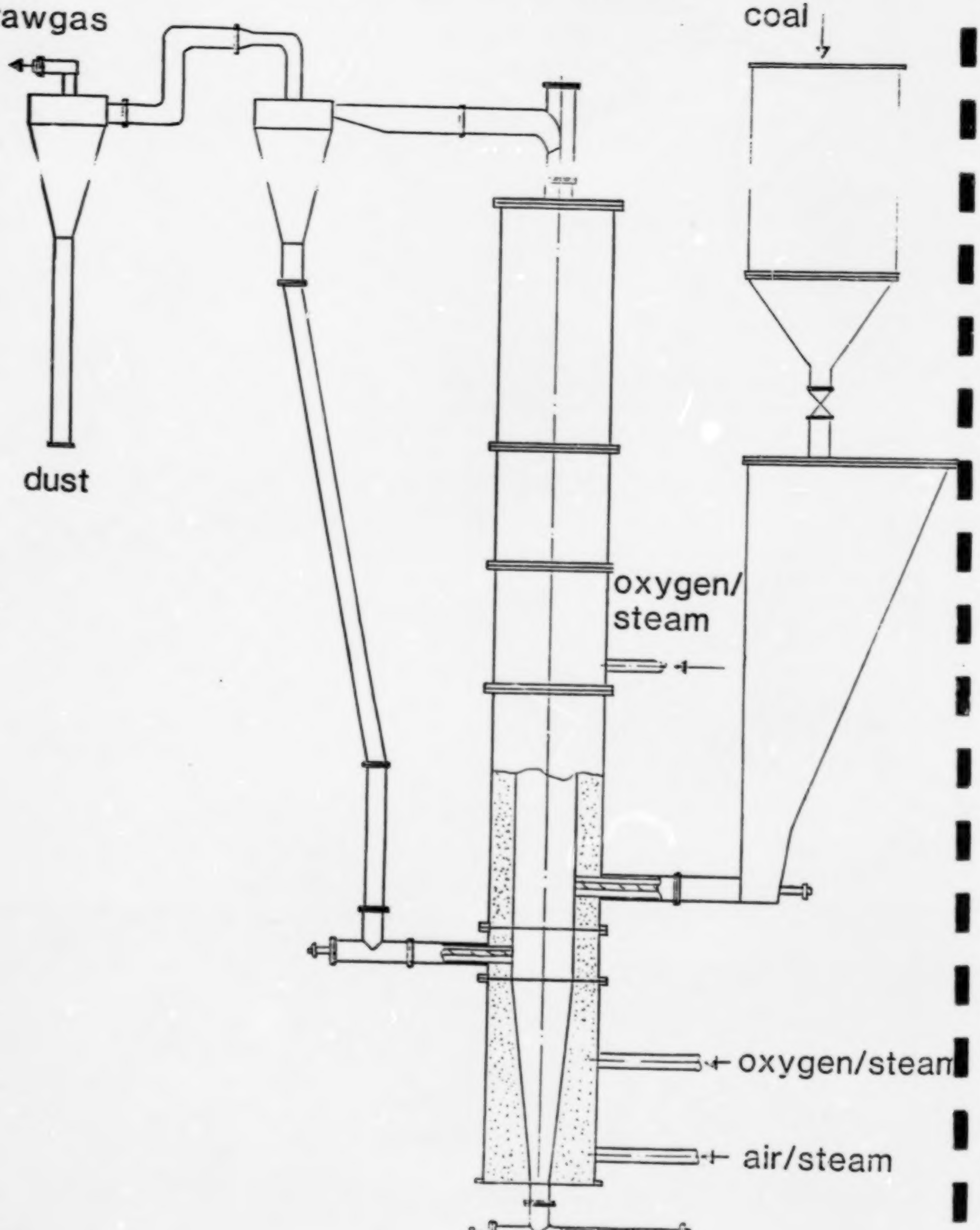
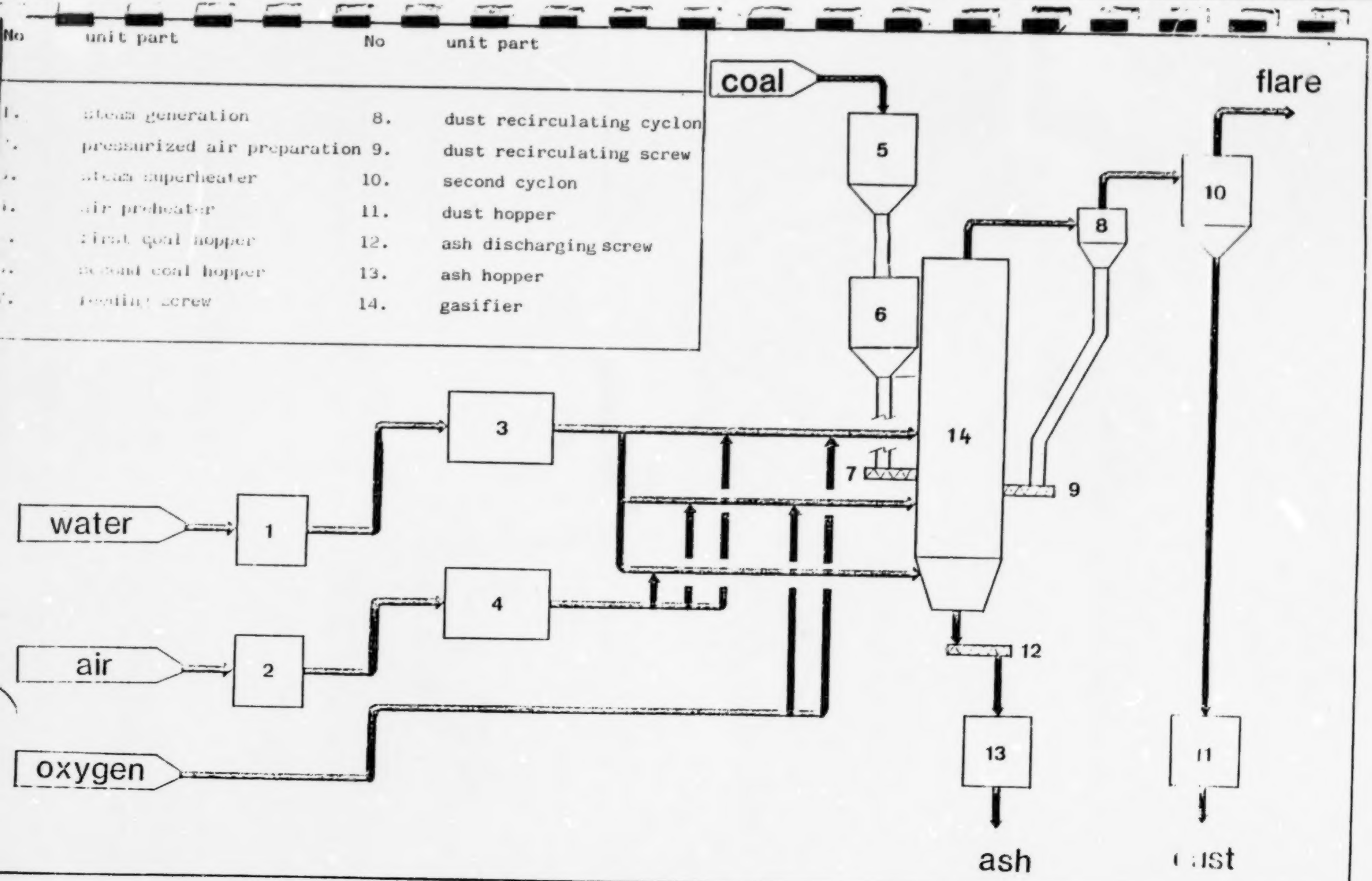

.rawgas
coal
dust
oxygen/
steam
oxygen/stea
air/steam
ash

Fig.9

HTW-process development unit

RHEINBRAUN


coal
flare
5
6
10
8
14
7
9
3
1
4
2
12
13
11
water
air
oxygen
ash
dust

No	unit part	No	unit part
1.	steam generation	8.	dust recirculating cyclon
.	pressurized air preparation	9.	dust recirculating screw
.	steam superheater	10.	second cyclon
.	air preheater	11.	dust hopper
.	first coal hopper	12.	ash discharging screw
.	second coal hopper	13.	ash hopper
.	feeding screw	14.	gasifier

Fig.10

HTW-process development unit flowsheet

RHEINBR UN

Fig.12 **gasification tests with Ontario-lignite**

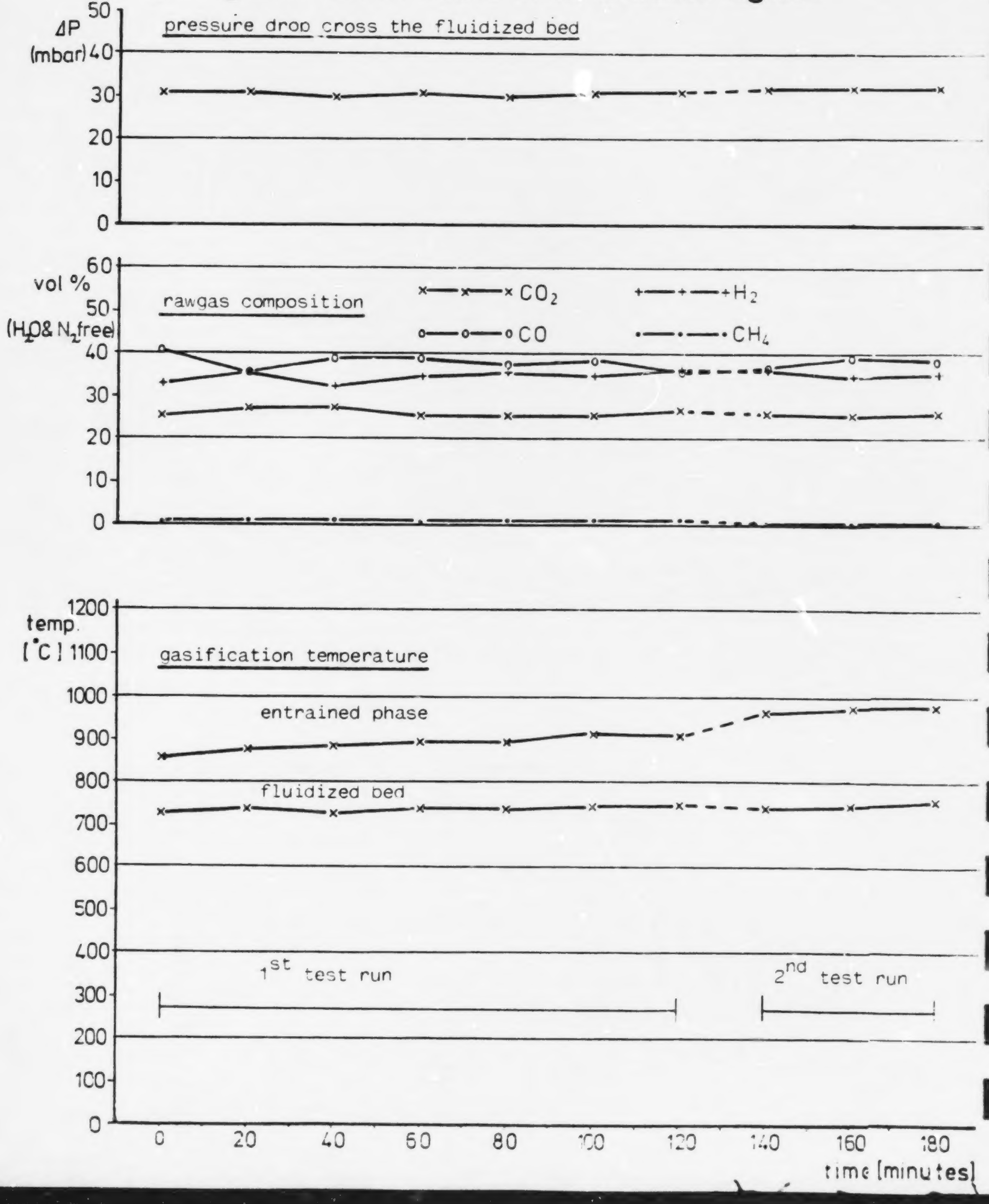


Fig. 15: gasification test with Saskatchewan-lignite

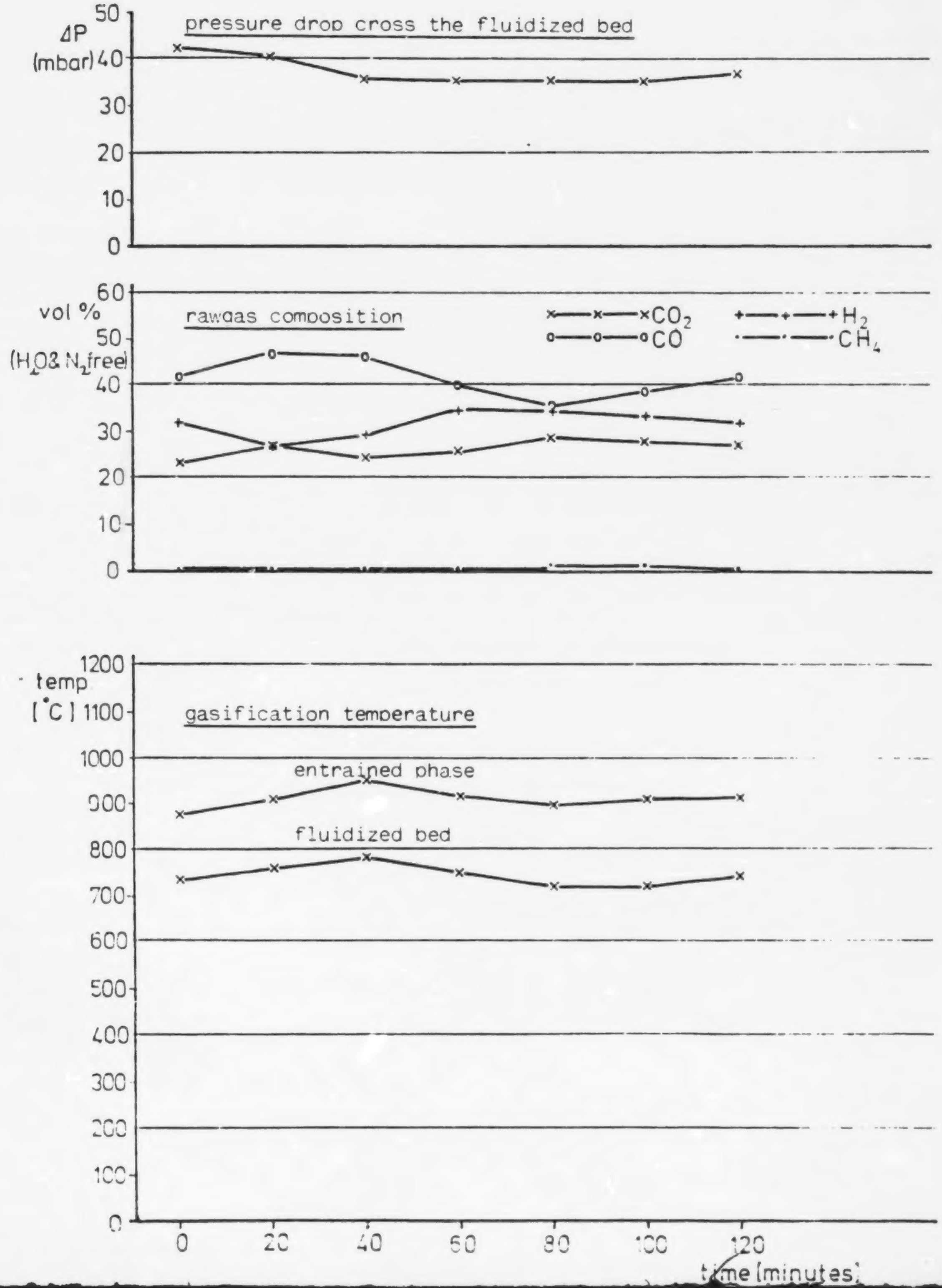


Table 1 Lignite analysis done by OEC

			Ontario Lignite		Saskatchewan Lignite
			Sample 1	Sample 2	
Elementary analysis					
(maf)		C	72.92	73.15	72.69
[% wt]		H	4.74	4.65	5.07
		O	20.00	19.31	20.19
		N	0.81	0.93	1.18
		S	1.53	1.96	0.87
Proximate analysis					
Moisture	(raw)	[% wt]	40-45	40-45	32.3
Ash	(mf)	[% wt]	16.79	20.82	15.8
Volatiles	(maf)	[% wt]	50.32	49.41	45.2
Ash analysis					
		SiO_2	39.97	39.11	43.9
[% wt]		Al_2O_3	12.01	11.45	17.7
		Fe_2O_3	9.63	10.77	4.31
		SO_3	14.93	16.20	9.24
		CaO	15.56	14.24	12.3
		MgO	4.29	3.68	2.54
		Na_2O	1.34	1.09	6.73
		K_2O	1.01	0.77	0.61
		MnO	-	-	0.006
		P_2O_5	0.13	0.16	0.54
		TiO_2	0.89	0.85	1.17

Table 2

Screen Analysis, Density and Pore Volume

Grain Size (µm)	Ontario-Lignite	Saskatchewan-Lignite
> 4,000	-	0.06
4,000 - 3,150	-	0.51
3,150 - 2,000	6.00	17.23
2,000 - 1,000	24.20	36.92
1,000 - 500	15.92	21.92
500 - 315	10.00	9.48
315 - 200	8.12	5.26
200 - 125	7.44	3.59
125 - 90	4.88	1.71
90 - 63	5.64	1.33
< 63	17.28	1.99
average grain diameter (µm)	136	521
bulk density (g/cm^3)	0.72	0.72
apparent density (g/cm^3)	1.02	1.22
true density (g/cm^3)	1.64	1.56
pore volume (%)	38.00	22.26

Table 3

Comparison of Characteristical Data on Gasification of Ontario-Lignite and Saskatchewan-Lignite

		Ontario-Lignite	Saskatchewan-Lignite
Elementary Analysis			
	C	72.4	71.6
(% wt)	H	4.4	4.7
(maf)	O	20.4	21.8
	N	0.8	1.1
	S	2.0	0.8
Proximate Analysis			
	fixed carbon	36.1	46.8
(% wt) (mf)	volatile	36.0	38.8
	ash	27.9	14.4
Heating Value (waf) (KJ/kg)		25,870	26,890
Ignition Temperature			
(^{o}C)	coal	195	195
	coke	310	285
Reaction Velocity Constant k_m			
($cm^3/g\cdot s$)	850^{o}C	1.46	1.88
	900^{o}C	3.29	4.42
Ash Fusion Behaviour (Reducing Atmosphere)			
	sintering point	1,090	1,090
(^{o}C)	softening point	1,260	1,210
	hemispherical point	1,280	1,260
	flow point	1,340	1,310

Table 5: Results of Gasification tests

test setting		Ontario-lignite				Saskatchewan-	
No.		1		2		1	
coal feed rate	[kg $coal_{maf}$/h]	17.1		18.4		16.	
temperature	[°C]						
(fluidized-bed)		740		750		755	
(entrained phase)		895		975		915	
nitrogen**	$\left[\frac{m^3 (STP)}{kg\ coal_{maf}}\right]$	0.23		0.22		0.25	
oxygen**	$\left[\frac{m^3 (STP)}{kg\ coal_{maf}}\right]$	0.50		0.57		0.55	
steam	[kg/kg $coal_{maf}$]	0.41		0.38		0.43	
oxygen/steam	[m^3/kg]	1.36		1.50		1.29	
C-conversion rate	[%]	85.8		88.6		77.8	
specivic synthesis gas yield	$\left[\frac{m^3 (STP) CO+H_2}{t\ coal_{maf}}\right]$	1280		1350		1100	
raw gas analysis	[vol%]						
	CO	25.9	33.0	27.1	33.5	24.4	34.7
	H_2	23.7	30.2	25.4	31.4	18.7	26.6
	CH_4	0.6	0.8	0.3	0.4	0.3	0.5
	CO_2	17.7	22.6	18.3	22.6	15.7	22.3
	H_2O	21.6	-	19.1	-	29.7	-
	N_2	10.5	13.4	9.8	12.1	11.2	15.9
raw gas (H_2O and N_2 free)	[vol%]						
	CO	38.1		38.7		41.3	
	H_2	34.9		35.7		31.6	
	CH_4	0.9		0.5		0.6	
	CO_2	26.1		25.7		26.5	

** calculated from air as gasification agent, oxygen in air is added to oxygen supply

APPENDIX 3

DISCOUNTED CASH FLOWS AND FINANCIAL ANALYSIS COMPUTER PRINTOUTS

DCF ANALYSIS

LIGNITE TO METHANOL

METHANOL ASSESSMENT PROGRAM

[illegible]UNTED [illegible] FLOW A[illegible]S

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	263.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	263.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					79.3	88.5	95.0	96.0	97.0	98.0	99.1	100.1	101.2	102.3	103.4	104.5	105.6	106.8	107.9	109.1	110.3	111.6	112.8	114.1	
Total Revenues					133.2	162.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.8	208.1	211.4	214.8	218.2	221.7	225.3	228.9	232.5	236.3	240.0	
Gross Profit					53.9	73.9	88.3	90.2	92.2	94.2	96.2	98.3	100.4	102.6	104.7	106.9	109.2	111.5	113.8	116.2	118.5	121.0	123.5	126.0	
Cash Flow	(87.7)	(263.1)	(307.0)	(231.2)	53.9	73.9	88.3	90.2	92.2	94.2	96.2	98.3	100.4	102.6	104.7	106.9	109.2	111.5	113.8	116.2	118.5	121.0	123.5	126.0	1,152.3

INTERNAL RATE OF RETURN
(npv=0)

IRR	7.5

CASE: Base Case

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	67.7	283.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	67.7	283.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					79.3	88.5	95.0	96.0	97.0	98.0	99.1	100.1	101.2	102.3	103.4	104.5	105.6	106.8	107.9	109.1	110.3	111.6	112.8	114.1	
Total Revenues					146.8	179.0	202.0	205.3	208.5	211.9	215.3	218.7	222.2	225.8	229.4	233.0	236.8	240.6	244.4	248.3	252.3	256.3	260.4	264.6	
Gross Profit					67.5	90.5	107.0	109.2	111.5	113.8	116.2	118.6	121.0	123.5	126.0	128.6	131.2	133.8	136.5	139.2	142.0	144.8	147.6	150.5	
Cash Flow	(67.7)	(283.1)	(307.0)	(231.2)	67.5	90.5	107.0	109.2	111.5	113.8	116.2	118.6	121.0	123.5	126.0	128.6	131.2	133.8	136.5	139.2	142.0	144.8	147.6	150.5	1,570.0

INTERNAL RATE OF RETURN (npv=0)	
IRR	9.5

CASE: Sensitivity Case A+

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 19.4 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Methanol selling price increased by 10%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	263.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	263.1	307.0	231.3																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.9	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.9	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					79.3	88.5	95.0	96.0	97.0	98.0	99.1	100.1	101.2	102.3	103.4	104.5	105.8	106.8	107.9	109.1	110.3	111.6	112.8	114.1	
Total Revenues					121.1	147.6	166.8	169.3	172.0	174.7	177.5	180.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.8	208.1	211.4	214.8	218.2	
Gross Profit					41.8	59.1	71.8	73.3	75.0	76.7	78.5	80.3	82.1	83.9	85.8	87.7	89.7	91.6	93.6	95.7	97.7	99.8	102.0	104.2	
Cash Flow	(87.7)	(263.1)	(307.0)	(231.3)	41.8	59.1	71.8	73.3	75.0	76.7	78.5	80.3	82.1	83.9	85.8	87.7	89.7	91.6	93.6	95.7	97.7	99.8	102.0	104.2	781.1

INTERNAL RATE OF RETURN
(npv=0)

IRR	5.5

CASE: Sensitivity Case A-

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 16.0 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Methanol selling price decreased by 10%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

METHANOL [illegible] PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	283.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	283.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					32.9	39.6	44.2	44.4	44.6	44.7	44.9	45.1	45.3	45.5	45.6	45.8	46.0	46.2	46.4	46.6	46.7	46.9	47.1	47.3	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					32.9	39.6	44.2	44.4	44.6	44.7	44.9	45.1	45.3	45.5	45.6	45.8	46.0	46.2	46.4	46.6	46.7	46.9	47.1	47.3	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					73.5	81.5	87.2	88.1	89.0	90.0	90.9	91.9	92.9	93.9	94.9	95.9	97.0	98.0	99.1	100.2	101.3	102.5	103.6	104.8	
Total Revenues					133.2	162.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.8	208.1	211.4	214.8	218.2	221.7	225.3	228.9	232.5	236.3	240.0	
Gross Profit					59.6	80.8	96.1	98.1	100.1	102.2	104.4	106.5	108.7	110.9	113.2	115.5	117.8	120.2	122.6	125.1	127.6	130.1	132.7	135.3	
Cash Flow	(87.7)	(283.1)	(307.0)	(231.2)	59.6	80.8	96.1	98.1	100.1	102.2	104.4	106.5	108.7	110.9	113.2	115.5	117.8	120.2	122.6	125.1	127.6	130.1	132.7	135.3	1,318.5

INTERNAL RATE OF RETURN
(npv=0)

IRR	8.3

CASE: Sensitivity Case 8+

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $13.8/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.8 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .4%
- methanol selling price 1.8%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	263.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	263.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					44.6	53.8	60.1	60.5	60.9	61.2	61.6	62.0	62.3	62.7	63.1	63.5	63.8	64.2	64.6	65.0	65.4	65.8	66.2	66.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					44.6	53.8	60.1	60.5	60.9	61.2	61.6	62.0	62.3	62.7	63.1	63.5	63.8	64.2	64.6	65.0	65.4	65.8	66.2	66.6	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					85.2	95.7	103.2	104.2	105.4	106.5	107.6	108.8	109.9	111.1	112.3	113.6	114.8	116.1	117.4	118.7	120.0	121.3	122.7	124.1	
Total Revenues					133.2	162.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.8	208.1	211.4	214.8	218.2	221.7	225.3	228.9	232.5	236.3	240.0	
Gross Profit					47.9	66.6	80.1	82.0	83.8	85.7	87.7	89.6	91.6	93.7	95.8	97.9	100.0	102.2	104.4	106.6	108.9	111.2	113.6	116.0	
Cash Flow	(87.7)	(263.1)	(307.0)	(231.2)	47.9	66.6	80.1	82.0	83.8	85.7	87.7	89.6	91.6	93.7	95.8	97.9	100.0	102.2	104.4	106.6	108.9	111.2	113.6	116.0	97[illegible].4

INTERNAL RATE OF RETURN
(npv=0)

IRR	6.5

CASE: Sensitivity Case B-

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $18.7/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Escalation factors used for:
- manpower 2%
- maintenunce 2%
- utilities 1%
- tax and insurance .5%
- lignite price .6%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	263.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	263.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.6	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					10.9	11.1	11.3	11.6	11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	
Maintenance					15.5	15.8	16.1	16.4	16.8	17.1	17.5	17.8	18.2	18.5	18.9	19.3	19.7	20.1	20.5	20.9	21.3	21.7	22.1	22.6	
Utilities					2.4	2.9	3.3	3.3	3.3	3.4	3.4	3.4	3.5	3.5	3.5	3.6	3.6	3.6	3.7	3.7	3.8	3.8	3.8	3.9	
Taxes and Insurance					5.1	5.1	5.2	5.2	5.2	5.2	5.3	5.3	5.3	5.3	5.4	5.4	5.4	5.4	5.5	5.5	5.5	5.6	5.6	5.6	
Total Operating Costs					33.9	35.0	35.9	36.5	37.1	37.7	38.4	39.0	39.7	40.4	41.1	41.8	42.5	43.2	44.0	44.7	45.5	46.3	47.1	47.9	
CASHFLOW																									
Total Expenses					72.5	81.5	87.9	88.7	89.6	90.5	91.4	92.3	93.3	94.2	95.2	96.2	97.2	98.2	99.2	100.2	101.3	102.3	103.4	104.5	
Total Revenues					133.2	162.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.8	208.1	211.4	214.8	218.2	221.7	225.3	228.9	232.5	236.3	240.0	
Gross Profit					60.7	80.8	95.4	97.5	99.6	101.7	103.9	106.1	108.3	110.6	112.9	115.3	117.6	120.1	122.5	125.1	127.6	130.2	132.8	135.5	
Cash Flow	(87.7)	(263.1)	(307.0)	(231.2)	60.7	80.8	95.4	97.5	99.6	101.7	103.9	106.1	108.3	110.6	112.9	115.3	117.6	120.1	122.5	125.1	127.6	130.2	132.8	135.5	1,315.1

INTERNAL RATE OF RETURN
(npv=0)

IRR	8.3

CASE: Sensitivity Case C+

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $18.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)
Total operating costs decreased by 20%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	263.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	263.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					15.7	16.0	16.3	16.7	17.0	17.3	17.7	18.0	18.4	18.8	19.1	19.5	19.9	20.3	20.7	21.1	21.6	22.0	22.4	22.9	
Maintenance					22.3	22.7	23.2	23.7	24.1	24.6	25.1	25.6	26.1	26.7	27.2	27.7	28.3	28.8	29.4	30.0	30.6	31.2	31.8	32.5	
Utilities					3.4	4.2	4.7	4.7	4.8	4.8	4.9	4.9	5.0	5.0	5.1	5.1	5.2	5.2	5.3	5.3	5.4	5.4	5.5	5.6	
Taxes and Insurance					7.3	7.3	7.4	7.4	7.4	7.5	7.5	7.6	7.6	7.6	7.7	7.7	7.8	7.8	7.8	7.9	7.9	7.9	8.0	8.0	
Total Operating Costs					48.7	50.3	51.6	52.5	53.4	54.3	55.2	56.1	57.1	58.1	59.1	60.1	61.1	62.2	63.3	64.4	65.5	66.6	67.8	68.9	
CASHFLOW																									
Total Expenses					87.4	96.8	103.6	104.7	105.9	107.1	108.2	109.4	110.7	111.9	113.2	114.5	115.8	117.1	118.5	119.8	121.2	122.6	124.1	125.5	
Total Revenues					133.2	162.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.8	208.1	211.4	214.8	218.2	221.7	225.3	228.9	232.5	236.3	240.0	
Gross Profit					45.8	65.5	79.7	81.5	83.3	85.2	87.0	89.0	90.9	92.9	94.9	96.9	99.0	101.1	103.3	105.5	107.7	109.9	112.2	114.5	
Cash Flow	(87.7)	(263.1)	(307.0)	(231.2)	45.8	65.5	79.7	81.5	83.3	85.2	87.0	89.0	90.9	92.9	94.9	96.9	99.0	101.1	103.3	105.5	107.7	109.9	112.2	114.5	956.8

INTERNAL RATE OF RETURN
(npv=0)

IRR	6.4

CASE: Sensitivity Case C -

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.8 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)
Total operating costs increased by 20%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM
DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	79.7	239.1	279.0	199.3																					
Working Capital				12.0																					
Total Investment	79.7	239.1	279.0	211.2																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.9	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.9	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					79.3	88.5	95.0	96.0	97.0	98.0	99.1	100.1	101.2	102.3	103.4	104.5	105.6	106.8	107.9	109.1	110.3	111.6	112.8	114.1	
Total Revenues					133.2	162.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.9	208.1	211.4	214.8	218.2	221.7	225.3	228.9	232.5	236.3	240.0	
Gross Profit					53.9	73.9	88.3	90.2	92.2	94.2	[illegible].2	98.3	100.4	102.6	104.7	106.9	109.2	111.5	113.8	116.2	118.5	121.0	123.5	126.0	
Cash Flow	(79.7)	(239.1)	(279.0)	(211.2)	53.9	73.9	88.3	90.2	92.2	94.2	96.2	98.3	100.4	102.6	104.7	106.9	109.2	111.5	113.8	116.2	118.5	121.0	123.5	126.0	1,232.3

INTERNAL RATE OF RETURN
(npv=0)

IRR	8.5

CASE: Sensitivity Case D+

Total plant cost: $797 MM
Working capital: $12MM
Total investment: $809 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Total plant costs decreased by 10%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	96.4	289.3	337.5	241.1																					
Working Capital				12.0																					
Total Investment	96.4	289.3	337.5	253.1																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.9	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.9	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					79.3	88.5	95.0	96.0	97.0	98.0	99.1	100.1	101.2	102.3	103.4	104.5	105.6	106.8	107.9	109.1	110.3	111.6	112.8	114.1	
Total Revenues					133.2	162.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.8	208.1	211.4	214.8	218.2	221.7	225.3	228.9	232.5	236.3	240.0	
Gross Profit					53.9	73.9	88.3	90.2	92.2	94.2	96.2	98.3	100.4	102.6	104.7	106.9	109.2	111.5	113.8	116.2	118.5	121.0	123.5	126.0	
Cash Flow	(96.4)	(289.3)	(337.5)	(253.1)	53.9	73.9	88.3	90.2	92.2	94.2	96.2	98.3	100.4	102.6	104.7	106.9	109.2	111.5	113.8	116.2	118.5	121.0	123.5	126.0	1,064.9

INTERNAL RATE OF RETURN
(npv=0)

IRR	6.5

CASE: Sensitivity Case D-

Total plant cost: $964.4 MM
Working capital: $12MM
Total investment: $976.4 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Total plant costs increased by 10%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTR
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	283.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	283.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					13.1	13.4	13.8	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.8	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.6	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					79.3	88.5	95.0	96.0	97.0	98.0	99.1	100.1	101.2	102.3	103.4	104.5	105.6	106.8	107.9	109.1	110.3	111.6	112.8	114.1	
Total Revenues					133.2	162.6	183.9	187.1	190.4	193.7	197.1	200.6	204.1	207.7	211.4	215.1	218.9	222.8	226.7	230.7	234.7	238.9	243.1	247.3	
Gross Profit					53.9	74.1	88.8	91.1	93.4	95.7	98.1	100.5	103.0	105.5	108.0	110.6	113.3	116.0	118.7	121.5	124.4	127.3	130.2	133.3	
Cash Flow	(87.7)	(283.1)	(307.0)	(231.2)	53.9	74.1	88.8	91.1	93.4	95.7	98.1	100.5	103.0	105.5	108.0	110.6	113.3	116.0	118.7	121.5	124.4	127.3	130.2	133.3	1,218.4

INTERNAL RATE OF RETURN
(npv=0)

IRR	7.7

CASE: Sensitivity Case E+

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.78%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	263.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	263.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					79.3	88.5	95.0	96.0	97.0	98.0	99.1	100.1	101.2	102.3	103.4	104.5	105.6	106.8	107.9	109.1	110.3	111.6	112.8	114.1	
Total Revenues					133.2	162.1	182.7	185.4	188.1	190.8	193.6	196.4	199.2	202.1	205.0	208.0	211.0	214.1	217.2	220.3	223.5	226.8	230.1	233.4	
Gross Profit					53.9	73.6	87.7	89.4	91.1	92.8	94.5	96.3	98.0	99.8	101.7	103.5	105.4	107.3	109.3	111.2	113.2	115.2	117.3	119.3	
Cash Flow	(87.7)	(263.1)	(307.0)	(231.2)	53.9	73.5	87.7	89.4	91.1	92.8	94.5	96.3	98.0	99.8	101.7	103.5	105.4	107.3	109.3	111.2	113.2	115.2	117.3	119.3	1,091.5

INTERNAL RATE OF RETURN
(npv=0)

IRR	7.2

CASE: Sensitivity Case E-

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.45%

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	263.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	263.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					39.8	48.0	53.6	53.9	54.2	54.4	54.7	55.0	55.2	55.5	55.8	56.1	56.4	56.6	56.9	57.2	57.5	57.8	58.1	58.4	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					39.8	48.0	53.6	53.9	54.2	54.4	54.7	55.0	55.2	55.5	55.8	56.1	56.4	56.6	56.9	57.2	57.5	57.8	58.1	58.4	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					80.5	89.9	96.6	97.6	98.6	99.7	100.7	101.8	102.9	103.9	105.1	106.2	107.3	108.5	109.7	110.9	112.1	113.3	114.6	115.8	
Total Revenues					137.3	167.4	189.0	192.0	195.1	198.2	201.4	204.6	207.9	211.2	214.6	218.0	221.5	225.1	228.7	232.3	236.0	239.8	243.7	247.5	
Gross Profit					56.8	77.5	92.4	94.4	96.5	98.5	100.7	102.8	105.0	107.3	109.5	111.8	114.2	116.6	119.0	121.5	124.0	126.5	129.1	131.7	
Cash Flow	(87.7)	(263.1)	(307.0)	(231.2)	56.8	77.5	92.4	94.4	96.5	98.5	100.7	102.8	105.0	107.3	109.5	111.8	114.2	116.6	119.0	121.5	124.0	126.5	129.1	131.7	1,246.8

INTERNAL RATE OF RETURN
(npv=0)

IRR	7.9

CASE: Sensitivity Case F+

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	87.7	263.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	87.7	263.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					37.4	45.1	50.4	50.6	50.9	51.1	51.4	51.6	51.9	52.2	52.4	52.7	52.9	53.2	53.5	53.7	54.0	54.3	54.6	54.8	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					37.4	45.1	50.4	50.6	50.9	51.1	51.4	51.6	51.9	52.2	52.4	52.7	52.9	53.2	53.5	53.7	54.0	54.3	54.6	54.8	
OPERATING COSTS																									
Manpower					13.1	13.4	13.6	13.9	14.2	14.5	14.8	15.0	15.3	15.7	16.0	16.3	16.6	16.9	17.3	17.6	18.0	18.3	18.7	19.1	
Maintenance					18.6	19.0	19.4	19.7	20.1	20.5	20.9	21.4	21.8	22.2	22.7	23.1	23.6	24.1	24.5	25.0	25.5	26.0	26.6	27.1	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.9	43.0	43.7	44.5	45.2	46.0	46.8	47.6	48.4	49.3	50.1	51.0	51.8	52.7	53.6	54.6	55.5	56.5	57.5	
CASHFLOW																									
Total Expenses					78.1	87.0	93.4	94.4	95.4	96.4	97.4	98.4	99.5	100.6	101.7	102.8	103.9	105.1	106.2	107.4	108.6	109.8	111.0	112.3	
Total Revenues					129.0	157.3	177.5	180.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.8	208.1	211.4	214.8	218.2	221.7	225.3	228.9	232.5	
Gross Profit					50.9	70.3	84.2	86.0	87.9	89.8	91.8	93.8	95.8	97.8	99.9	102.0	104.2	106.4	108.6	110.8	113.1	115.5	117.8	120.2	
Cash Flow	(87.7)	(263.1)	(307.0)	(231.2)	50.9	70.3	84.2	86.0	87.9	89.8	91.8	93.8	95.8	97.8	99.9	102.0	104.2	106.4	108.6	110.8	113.1	115.5	117.8	120.2	1,057.9

INTERNAL RATE OF RETURN
(npv=0)

IRR	7.0

CASE: Sensitivity Case F-

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 310
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	67.7	283.1	307.0	219.3																					
Working Capital				12.0																					
Total Investment	67.7	283.1	307.0	231.2																					
FEEDSTOCKS																									
Lignite					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
Wood																									
Peat																									
Natural Gas																									
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					38.6	46.6	52.0	52.3	52.5	52.8	53.0	53.3	53.6	53.8	54.1	54.4	54.7	54.9	55.2	55.5	55.8	56.0	56.3	56.6	
OPERATING COSTS																									
Manpower					13.1	13.2	13.4	13.5	13.6	13.8	13.9	14.0	14.2	14.3	14.5	14.6	14.8	14.9	15.1	15.2	15.4	15.5	15.7	15.8	
Maintenance					18.6	18.8	19.0	19.2	19.4	19.5	19.7	19.9	20.1	20.3	20.5	20.8	21.0	21.2	21.4	21.6	21.8	22.0	22.2	22.5	
Utilities					2.8	3.5	3.9	3.9	4.0	4.0	4.0	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.5	4.6	
Taxes and Insurance					6.1	6.1	6.2	6.2	6.2	6.3	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	
Total Operating Costs					40.7	41.6	42.4	42.8	43.2	43.6	44.0	44.4	44.8	45.2	45.6	46.1	46.5	46.9	47.3	47.8	48.2	48.7	49.1	49.6	
CASHFLOW																									
Total Expenses					79.3	88.2	94.4	95.0	95.7	96.3	97.0	97.7	98.4	99.0	99.7	100.4	101.1	101.8	102.5	103.3	104.0	104.7	105.5	106.2	
Total Revenues					133.2	162.4	183.3	186.2	189.2	192.2	195.3	198.4	201.6	204.8	208.1	211.4	214.8	218.2	221.7	225.3	228.9	232.5	236.3	240.0	
Gross Profit					53.9	74.2	88.9	91.2	93.5	95.9	98.3	100.7	103.2	105.8	108.4	111.0	113.7	116.4	119.2	122.0	124.9	127.8	130.8	133.9	
Cash Flow	(67.7)	(283.1)	(307.0)	(231.2)	53.9	74.2	88.9	91.2	93.5	95.9	98.3	100.7	103.2	105.8	108.4	111.0	113.7	116.4	119.2	122.0	124.9	127.8	130.8	133.9	1,224.6

INTERNAL RATE OF RETURN
(npv=0)

IRR 7.8

CASE: Sensitivity Case G

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.2/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 17.6 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Escalation factors used for:
- manpower 1%
- maintenance 1%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	28.7	86.1	100.4	71.7																					
Working Capital				23.0																					
Total Investment	28.7	86.1	100.4	94.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					139.8	173.0	176.8	180.7	184.6	188.6	192.8	197.0	201.3	205.7	210.2	214.8	219.6	224.4	229.3	234.3	239.5	244.8	250.1	255.6	
Total Revenues					159.8	202.9	206.2	209.5	212.8	216.2	219.7	223.2	226.8	230.4	234.1	237.8	241.7	245.5	249.4	253.4	257.5	261.6	265.8	270.1	
Gross Profit					20.0	29.9	29.4	28.8	28.2	27.6	26.9	26.2	25.5	24.7	23.9	23.0	22.1	21.1	20.1	19.1	18.0	16.9	15.7	14.4	
Cash Flow	(28.7)	(86.1)	(100.4)	(94.8)	20.0	29.9	29.4	28.8	28.2	27.6	26.9	26.2	25.5	24.7	23.9	23.0	22.1	21.1	20.1	19.1	18.0	16.9	15.7	14.4	151.8

INTERNAL RATE OF RETURN
(npv=0)

IRR	4.0

CASE: Sensitivity Case H

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Net gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.6%

GAS TO METHANOL (2,500 TPD)

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					139.8	173.0	176.8	180.7	184.6	188.6	192.8	197.0	201.3	205.7	210.2	214.8	219.6	224.4	229.3	234.3	239.5	244.8	250.1	255.6	
Total Revenues					159.8	202.9	206.2	209.5	212.8	216.2	219.7	223.2	226.8	230.4	234.1	237.8	241.7	245.5	249.4	253.4	257.5	261.6	265.8	270.1	
Gross Profit					20.0	29.9	29.4	28.8	28.2	27.6	26.9	26.2	25.5	24.7	23.9	23.0	22.1	21.1	20.1	19.1	18.0	16.8	15.7	14.4	
Cash Flow		(57.4)	(114.8)	(137.8)	20.0	29.9	29.4	28.8	28.2	27.6	26.9	26.2	25.5	24.7	23.9	23.0	22.1	21.1	20.1	19.1	18.0	16.8	15.7	14.4	151.6

INTERNAL RATE OF RETURN
(npv=0)

IRR	4.2

CASE: Base Case

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Net gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.6%

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					139.8	173.0	176.8	180.7	184.6	188.6	192.8	197.0	201.3	205.7	210.2	214.8	219.6	224.4	229.3	234.3	239.5	244.8	250.1	255.6	
Total Revenues					175.6	223.0	226.6	230.2	233.9	237.6	241.4	245.3	249.2	253.2	257.3	261.4	265.6	269.8	274.1	278.5	283.0	287.5	292.1	296.8	
Gross Profit					35.8	50.0	49.8	49.5	49.3	49.0	48.7	48.3	47.9	47.5	47.0	46.5	46.0	45.4	44.8	44.2	43.5	42.7	42.0	41.1	
Cash Flow		(57.4)	[illegible]14.8)	(137.8)	35.8	50.0	49.8	49.5	49.3	49.0	48.7	48.3	47.9	47.5	47.0	46.5	46.0	45.4	44.8	44.2	43.5	42.7	42.0	41.1	609.2

INTERNAL RATE OF RETURN
(npv=0)

IRR	12.4

CASE: Sensitivity Case A+

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Net gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 21.1 cents/liter

Methanol selling price increased by 10%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.8%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM ®
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					139.8	173.0	176.8	180.7	184.6	188.6	192.8	197.0	201.3	205.7	210.2	214.8	219.6	224.4	229.3	234.3	239.5	244.8	250.1	255.6	
Total Revenues					145.6	185.0	187.9	190.9	194.0	197.1	200.2	203.5	206.7	210.0	213.4	216.8	220.3	223.8	227.4	231.0	234.7	238.5	242.3	246.1	
Gross Profit					5.8	12.0	11.1	10.3	9.4	8.5	7.5	6.5	5.4	4.3	3.1	1.9	0.7	(0.6)	(1.9)	(3.3)	(4.8)	(6.3)	(7.9)	(9.5)	
Cash Flow		(57.4)	(114.8)	(137.8)	5.8	12.0	11.1	10.3	9.4	8.5	7.5	6.5	5.4	4.3	3.1	1.9	0.7	(0.6)	(1.9)	(3.3)	(4.8)	(6.3)	(7.9)	(9.5)	(257.9)

INTERNAL RATE OF RETURN
(npv=0)

IRR ERR

CASE: Sensitivity Case A-

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Net gas yield: 0.0263 MSCF/liter
Initial selling price of methanol: 17.5 cents/liter

Methanol selling price decreased by 10%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					107.5	137.2	140.2	143.2	146.3	149.5	152.7	156.0	159.4	162.8	166.3	169.9	173.6	177.4	181.2	185.1	189.1	193.2	197.4	201.6	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					107.5	137.2	140.2	143.2	146.3	149.5	152.7	156.0	159.4	162.8	166.3	169.9	173.6	177.4	181.2	185.1	189.1	193.2	197.4	201.6	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					130.4	160.8	164.2	167.6	171.1	174.7	178.3	182.0	185.9	189.7	193.7	197.8	201.9	206.1	210.4	214.9	219.4	224.0	228.7	233.5	
Total Revenues					159.8	202.9	206.2	209.5	212.8	216.2	219.7	223.2	226.8	230.4	234.1	237.8	241.7	245.5	249.4	253.4	257.5	261.6	265.8	270.1	
Gross Profit					29.4	42.1	42.0	41.9	41.7	41.6	41.4	41.2	40.9	40.7	40.4	40.1	39.8	39.4	39.0	38.6	38.1	37.7	37.1	36.6	
Cash Flow		(57.4)	(114.8)	(137.8)	29.4	42.1	42.0	41.9	41.7	41.6	41.4	41.2	40.9	40.7	40.4	40.1	39.8	39.4	39.0	38.6	38.1	37.7	37.1	36.6	479.6

INTERNAL RATE OF RETURN
(npv=0)

IRR	10.1

CASE: Sensitivity Case 6+

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 4.91 $/MSCF
Nat gas yield: 0.0263 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.16%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					126.3	161.6	165.4	169.4	173.4	177.5	181.7	186.0	190.4	194.9	199.5	204.3	209.1	214.1	219.1	224.3	229.6	235.1	240.7	246.4	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					126.3	161.6	165.4	169.4	173.4	177.5	181.7	186.0	190.4	194.9	199.5	204.3	209.1	214.1	219.1	224.3	229.6	235.1	240.7	246.4	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					149.2	185.2	189.4	193.7	198.1	202.7	207.3	212.0	216.9	221.8	226.9	232.1	237.4	242.9	248.4	254.1	259.9	265.9	272.0	278.2	
Total Revenues					159.8	202.9	206.2	209.5	212.8	216.2	219.7	223.2	226.8	230.4	234.1	237.8	241.7	245.5	249.4	253.4	257.5	261.6	265.8	270.1	
Gross Profit					10.6	17.7	16.8	15.8	14.7	13.6	12.4	11.2	9.9	8.6	7.2	5.8	4.3	2.7	1.1	(0.6)	(2.4)	(4.2)	(6.2)	(8.1)	
Cash Flow		(57.4)	(114.8)	(137.8)	10.6	17.7	16.8	15.8	14.7	13.6	12.4	11.2	9.9	8.6	7.2	5.8	4.3	2.7	1.1	(0.6)	(2.4)	(4.2)	(6.2)	(8.1)	(178.4)

INTERNAL RATE OF RETURN
(npv=0)

IRR	ERR

CASE: Sensitivity Case 8-

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.77 $/MSCF
Net gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.37%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					4.6	4.7	4.8	4.9	5.0	5.1	5.2	5.3	5.4	5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	
Maintenance					9.8	10.0	10.2	10.4	10.6	10.8	11.0	11.3	11.5	11.7	11.9	12.2	12.4	12.7	12.9	13.2	13.5	13.7	14.0	14.3	
Utilities					0.9	1.1	1.1	1.1	1.1	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.2	1.3	1.3	1.3	1.3	1.3	1.3	1.3	
Taxes and Insurance					3.8	3.8	3.8	3.8	3.9	3.9	3.9	3.9	4.0	4.0	4.0	4.0	4.0	4.1	4.1	4.1	4.1	4.1	4.2	4.2	
Total Operating Costs					19.1	19.6	19.9	20.3	20.6	21.0	21.3	21.7	22.0	22.4	22.8	23.1	23.5	23.9	24.3	24.8	25.2	25.6	26.0	26.5	
CASHFLOW																									
Total Expenses					136.0	169.0	172.8	176.6	180.4	184.4	188.5	192.6	196.9	201.2	205.6	210.2	214.8	219.5	224.4	229.3	234.4	239.6	244.9	250.3	
Total Revenues					159.8	202.9	206.2	209.5	212.8	216.2	219.7	223.2	226.8	230.4	234.1	237.8	241.7	245.5	249.4	253.4	257.5	261.6	265.8	270.1	
Gross Profit					23.8	33.9	33.4	32.9	32.4	31.8	31.2	30.6	29.9	29.2	28.5	27.7	26.8	26.0	25.1	24.1	23.1	22.0	20.9	19.8	
Cash Flow		(57.4)	(114.8)	(137.8)	23.8	33.9	33.4	32.9	32.4	31.8	31.2	30.6	29.9	29.2	28.5	27.7	26.8	26.0	25.1	24.1	23.1	22.0	20.9	19.8	243.2

INTERNAL RATE OF RETURN
(npv=0)

IRR 6.2

CASE: Sensitivity Case C+

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Net gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter
Total operating costs decreased by 20%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					6.6	6.7	6.9	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	8.2	8.4	8.5	8.7	8.9	9.1	9.2	9.4	9.6	
Maintenance					14.2	14.5	14.8	15.1	15.4	15.7	16.0	16.3	16.6	17.0	17.3	17.7	18.0	18.4	18.7	19.1	19.5	19.9	20.3	20.7	
Utilities					1.3	1.6	1.6	1.6	1.7	1.7	1.7	1.7	1.7	1.7	1.8	1.8	1.8	1.8	1.8	1.9	1.9	1.9	1.9	1.9	
Taxes and Insurance					5.5	5.5	5.6	5.6	5.6	5.6	5.7	5.7	5.7	5.8	5.8	5.8	5.8	5.9	5.9	5.9	6.0	6.0	6.0	6.0	
Total Operating Costs					27.6	28.4	28.8	29.3	29.8	30.3	30.8	31.3	31.8	32.4	32.9	33.5	34.0	34.6	35.2	35.8	36.4	37.0	37.6	38.3	
CASHFLOW																									
Total Expenses					144.5	177.8	181.6	185.6	189.6	193.7	198.0	202.3	206.7	211.2	215.8	220.5	225.3	230.2	235.2	240.4	245.6	251.0	256.5	262.1	
Total Revenues					159.8	202.9	206.2	209.5	212.8	216.2	219.7	223.2	226.8	230.4	234.1	237.8	241.7	245.5	249.4	253.4	257.5	261.6	265.8	270.1	
Gross Profit					15.3	25.2	24.5	23.9	23.2	22.5	21.7	20.9	20.1	19.2	18.3	17.4	16.4	15.3	14.2	13.1	11.9	10.6	9.3	8.0	
Cash Flow		(57.4)	(114.8)	(137.8)	15.3	25.2	24.5	23.9	23.2	22.5	21.7	20.9	20.1	19.2	18.3	17.4	16.4	15.3	14.2	13.1	11.9	10.6	9.3	8.0	41.1

INTERNAL RATE OF RETURN
(npv=0)

IRR 1.3

CASE: Sensitivity Case C-

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Nat gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter
Total operating costs increased by 20%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		52.2	104.4	104.4																					
Working Capital				23.0																					
Total Investment		52.2	104.4	127.4																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.[illegible]	31.3	31.8	
CASHFLOW																									
Total Expenses					139.8	173.0	176.8	180.7	184.6	188.6	192.8	197.0	201.3	205.7	210.2	214.8	219.6	224.4	229.3	234.3	239.5	244.8	250.1	255.6	
Total Revenues					159.8	202.9	206.2	209.5	212.8	216.2	219.7	223.2	226.8	230.4	234.1	237.8	241.7	245.5	249.4	253.4	257.5	261.6	265.8	270.1	
Gross Profit					20.0	29.9	29.4	28.8	28.2	27.6	26.9	26.2	25.5	24.7	23.9	23.0	22.1	21.1	20.1	19.1	18.0	16.9	15.7	14.4	
Cash Flow		(52.2)	(104.4)	(127.4)	20.0	29.9	29.4	28.8	28.2	27.6	26.9	26.2	25.5	24.7	23.9	23.0	22.1	21.1	20.1	19.1	18.0	16.9	15.7	14.4	177.6

INTERNAL RATE OF RETURN
(npv=0)

IRR 5.2

CASE: Sensitivity Case D+

Total plant cost: $261 MM
Working capital: $23MM
Total investment: $284 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Nat gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Total plant costs decreased by 10%

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.6%

```
                                                              METHANOL ASSESSMENT PROGRAM

                                                            DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

                              1986   1987   1988   1989   1990   1991   1992   1993   1994   1995   1996   1997   1998   1999   2000   2001   2002   2003   2004   2005   2006   2007   2008   2009     TOTAL
                             -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----   -------

Total Plant Cost                     63.2  126.4  126.4
Working Capital                                    23.0
Total Investment                     63.2  126.4  149.4

FEEDSTOCKS
----------------------------
Lignite
Wood
Peat
Natural Gas                                              116.9  149.4  152.8  156.3  159.8  163.5  167.2  171.0  174.8  178.8  182.9  187.0  191.3  195.6  200.1  204.6  209.2  214.0  218.8  223.8
Electrolytic Hydrogen
    (Power Costs)
                             -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----
Total Feedstock Costs                                    116.9  149.4  152.8  156.3  159.8  163.5  167.2  171.0  174.8  178.8  182.9  187.0  191.3  195.6  200.1  204.6  209.2  214.0  218.8  223.8

OPERATING COSTS
----------------------------
Manpower                                                   5.5    5.6    5.7    5.8    6.0    6.1    6.2    6.3    6.4    6.6    6.7    6.8    7.0    7.1    7.3    7.4    7.6    7.7    7.9    8.0
Maintenance                                               11.8   12.0   12.3   12.5   12.8   13.0   13.3   13.6   13.8   14.1   14.4   14.7   15.0   15.3   15.6   15.9   16.2   16.5   16.9   17.2
Utilities                                                  1.0    1.3    1.3    1.3    1.4    1.4    1.4    1.4    1.4    1.4    1.4    1.5    1.5    1.5    1.5    1.5    1.5    1.5    1.6    1.6
Taxes and Insurance                                        4.6    4.6    4.6    4.7    4.7    4.7    4.7    4.8    4.8    4.8    4.8    4.9    4.9    4.9    4.9    5.0    5.0    5.0    5.0    5.1
                             -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----
Total Operating Costs                                     22.9   23.6   24.0   24.4   24.8   25.2   25.6   26.0   26.5   26.9   27.4   27.8   28.3   28.8   29.3   29.8   30.3   30.8   31.3   31.8

CASHFLOW
----------------------------
Total Expenses                                           139.8  173.0  176.8  180.7  184.6  188.6  192.8  197.0  201.3  205.7  210.2  214.8  219.6  224.4  229.3  234.3  239.5  244.8  250.1  255.6
Total Revenues                                           159.8  202.9  206.2  209.5  212.8  216.2  219.7  223.2  226.8  230.4  234.1  237.8  241.7  245.5  249.4  253.4  257.5  261.6  265.8  270.1
Gross Profit                                              20.0   29.9   29.4   28.8   28.2   27.6   26.9   26.2   25.5   24.7   23.9   23.0   22.1   21.1   20.1   19.1   18.0   16.9   15.7   14.4
                             -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----   -------
Cash Flow                          (63.2)(126.4)(149.4)   20.0   29.9   29.4   28.8   28.2   27.6   26.9   26.2   25.5   24.7   23.9   23.0   22.1   21.1   20.1   19.1   18.0   16.9   15.7   14.4     122.6

INTERNAL RATE OF RETURN
      (npv=0)
----------------------------
IRR                            3.2


CASE:  Sensitivity Case D-

Total plant cost: $316 MM                                   Initial price of natural gas: 5.34 $/MSCF                     Escalation factors used for:
Working capital: $23MM                                      Nat gas yield: 0.0263 MSCF/liter                                - manpower 2%
Total investment: $339 MM                                   Initial selling price of methanol: 19.2 cents/liter             - maintenance 2%
Plant capacity: 2500 MTPD                                                                                                   - utilities 1%
On-stream days per year: 330                                Total plant costs increased by 10%                              - tax and insurance .5%
Plant utilization factor: 80% initial year                                                                                  - nat.gas price 2.27%
                          100% onward                                                                                       - methanol selling price 1.6%
```

METH[illegible] ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (40 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					139.8	173.0	176.8	180.7	184.6	188.6	192.8	197.0	201.3	205.7	210.2	214.8	219.6	224.4	229.3	234.3	239.5	244.8	250.1	255.6	
Total Revenues					159.8	203.3	206.9	210.5	214.2	217.9	221.8	225.7	229.7	233.7	237.8	242.0	246.3	250.6	255.0	259.5	264.1	268.7	273.4	278.2	
Gross Profit					20.0	30.3	30.1	29.8	29.6	29.3	29.0	28.7	28.4	28.0	27.6	27.2	26.7	26.2	25.7	25.2	24.6	24.0	23.3	22.6	
Cash Flow		(57.4)	(114.8)	(137.8)	20.0	30.3	30.1	29.8	29.6	29.3	29.0	28.7	28.4	28.0	27.6	27.2	26.7	26.2	25.7	25.2	24.6	24.0	23.3	22.6	226.0

INTERNAL RATE OF RETURN
(npv=0)

IRR	5.6

CASE: Sensitivity Case E+

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Net gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.76%

```
                                                      METHANOL ASSESSMENT PROGRAM

                                                      DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

                          1986   1987   1988   1989   1990   1991   1992   1993   1994   1995   1996   1997   1998   1999   2000   2001   2002   2003   2004   2005   2006   2007   2008   2009    TOTAL
                          -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -------

Total Plant Cost                  57.4  114.8  114.8
Working Capital                                 23.0
Total Investment                  57.4  114.8  137.8

FEEDSTOCKS
--------------------------

Lignite
Wood
Peat
Natural Gas                                           116.9  149.4  152.8  156.3  159.8  163.5  167.2  171.0  174.8  178.8  182.9  187.0  191.3  195.6  200.1  204.6  209.2  214.0  218.8  223.8
Electrolytic Hydrogen
    (Power Costs)
                          -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----
Total Feedstock Costs                                 116.9  149.4  152.8  156.3  159.8  163.5  167.2  171.0  174.8  178.8  182.9  187.0  191.3  195.6  200.1  204.6  209.2  214.0  218.8  223.8

OPERATING COSTS
--------------------------

Manpower                                                5.5    5.6    5.7    5.8    6.0    6.1    6.2    6.3    6.4    6.6    6.7    6.8    7.0    7.1    7.3    7.4    7.6    7.7    7.9    8.0
Maintenance                                            11.8   12.0   12.3   12.5   12.8   13.0   13.3   13.6   13.8   14.1   14.4   14.7   15.0   15.3   15.6   15.9   16.2   16.5   16.9   17.2
Utilities                                               1.0    1.3    1.3    1.3    1.4    1.4    1.4    1.4    1.4    1.4    1.4    1.5    1.5    1.5    1.5    1.5    1.5    1.5    1.6    1.6
Taxes and Insurance                                     4.6    4.8    4.6    4.7    4.7    4.7    4.7    4.8    4.8    4.8    4.8    4.9    4.9    4.9    4.9    5.0    5.0    5.0    5.0    5.1
                          -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----
Total Operating Costs                                  22.9   23.6   24.0   24.4   24.8   25.2   25.6   26.0   26.5   26.9   27.4   27.8   28.3   28.8   29.3   29.8   30.3   30.8   31.3   31.8

CASHFLOW
--------------------------

Total Expenses                                        139.8  173.0  176.8  180.7  184.6  188.6  192.8  197.0  201.3  205.7  210.2  214.8  219.6  224.4  229.3  234.3  239.5  244.8  250.1  255.6
Total Revenues                                        159.8  202.6  205.6  208.6  211.6  214.6  217.8  220.9  224.1  227.4  230.7  234.0  237.4  240.9  244.3  247.8  251.5  255.1  258.8  262.6
Gross Profit                                           20.0   29.6   28.8   27.9   27.0   26.0   25.0   23.9   22.8   21.7   20.4   19.2   17.8   16.5   15.0   13.5   12.0   10.4    8.7    6.9
                          -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -----  -------
Cash Flow                      (57.4)(114.8)(137.8)    20.0   29.6   28.8   27.9   27.0   26.0   25.0   23.9   22.8   21.7   20.4   19.2   17.8   16.5   15.0   13.5   12.0   10.4    8.7    6.9     83.2

INTERNAL RATE OF RETURN
       (npv=0)
--------------------------
IRR                        2.7


CASE:  Sensitivity Case E-

Total plant cost: $287 MM                                     Initial price of natural gas: 5.34 $/MSCF                     Escalation factors used for:
Working capital: $23MM                                        Net gas yield: 0.0283 MSCF/liter                                - manpower 2%
Total investment: $310 MM                                     Initial selling price of methanol: 19.2 cents/liter             - maintenance 2%
Plant capacity: 2500 MTPD                                                                                                     - utilities 1%
On-stream days per year: 330                                                                                                  - tax and insurance .5%
Plant utilization factor: 80% initial year                                                                                    - nat.gas price 2.27%
                          100% onward                                                                                         - methanol selling price 1.45%
```

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTR *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					120.4	153.9	157.4	161.0	164.7	168.4	172.2	176.1	180.1	184.2	188.4	192.7	197.1	201.5	206.1	210.8	215.6	220.5	225.5	230.6	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					120.4	153.9	157.4	161.0	164.7	168.4	172.2	176.1	180.1	184.2	188.4	192.7	197.1	201.5	206.1	210.8	215.6	220.5	225.5	230.6	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					143.4	177.5	181.4	185.4	189.4	193.6	197.8	202.2	206.6	211.1	215.8	220.5	225.4	230.3	235.4	240.5	245.8	251.2	256.8	262.4	
Total Revenues					164.6	209.1	212.4	215.8	219.3	222.8	226.4	230.0	233.7	237.4	241.2	245.1	249.0	253.0	257.0	261.1	265.3	269.5	273.9	278.2	
Gross Profit					21.3	31.6	31.0	30.4	29.8	29.2	28.5	27.8	27.1	26.3	25.4	24.5	23.6	22.7	21.6	20.6	19.5	18.3	17.1	15.8	
Cash Flow		(57.4)	(114.8)	(137.8)	21.3	31.6	31.0	30.4	29.8	29.2	28.5	27.8	27.1	26.3	25.4	24.5	23.6	22.7	21.6	20.6	19.5	18.3	17.1	15.8	182.1

INTERNAL RATE OF RETURN
(npv=0)

IRR	4.9

CASE: Sensitivity Case F+

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 340
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Nat gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					113.3	144.9	148.2	151.5	155.0	158.5	162.1	165.8	169.5	173.4	177.3	181.4	185.5	189.7	194.0	198.4	202.9	207.5	212.2	217.0	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					113.3	144.9	148.2	151.5	155.0	158.5	162.1	165.8	169.5	173.4	177.3	181.4	185.5	189.7	194.0	198.4	202.9	207.5	212.2	217.0	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					136.3	168.5	172.2	175.9	179.8	183.7	187.7	191.8	196.0	200.3	204.7	209.2	213.8	218.5	223.2	228.1	233.2	238.3	243.5	248.9	
Total Revenues					155.0	196.8	199.9	203.1	206.4	209.7	213.0	216.5	219.9	223.4	227.0	230.6	234.3	238.1	241.9	245.8	249.7	253.7	257.7	261.9	
Gross Profit					18.7	28.3	27.8	27.2	26.6	26.0	25.3	24.6	23.9	23.1	22.3	21.5	20.6	19.6	18.6	17.6	16.5	15.4	14.2	13.0	
Cash Flow		(57.4)	(114.8)	(137.8)	18.7	28.3	27.8	27.2	26.6	26.0	25.3	24.6	23.9	23.1	22.3	21.5	20.6	19.6	18.6	17.6	16.5	15.4	14.2	13.0	121.1

INTERNAL RATE OF RETURN
(npv=0)

IRR 3.5

CASE: Sensitivity Case F-

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Nat gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.6%

METHANOL ASSESSMENT PROGRAM
DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost		57.4	114.8	114.8																					
Working Capital				23.0																					
Total Investment		57.4	114.8	137.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					5.5	5.6	5.6	5.7	5.7	5.8	5.8	5.9	6.0	6.0	6.1	6.1	6.2	6.3	6.3	6.4	6.4	6.5	6.6	6.6	
Maintenance					11.8	11.9	12.0	12.2	12.3	12.4	12.5	12.7	12.8	12.9	13.0	13.2	13.3	13.4	13.6	13.7	13.8	14.0	14.1	14.3	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.8	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.4	23.6	23.8	24.0	24.3	24.5	24.7	24.9	25.2	25.4	25.6	25.8	26.1	26.3	26.6	26.8	27.0	27.3	27.5	
CASHFLOW																									
Total Expenses					139.8	172.8	176.4	180.1	183.8	187.7	191.7	195.7	199.8	204.0	208.3	212.6	217.1	221.7	226.4	231.1	236.0	241.0	246.1	251.3	
Total Revenues					159.8	202.9	206.2	209.5	212.8	216.2	219.7	223.2	226.8	230.4	234.1	237.8	241.7	245.5	249.4	253.4	257.5	261.6	265.8	270.1	
Gross Profit					20.0	30.1	29.8	29.4	29.0	28.5	28.0	27.5	27.0	26.4	25.8	25.2	24.5	23.8	23.1	22.3	21.5	20.6	19.7	18.7	
Cash Flow ·		(57.4)	(114.8)	(137.8)	20.0	30.1	29.8	29.4	29.0	28.5	28.0	27.5	27.0	26.4	25.8	25.2	24.5	23.8	23.1	22.3	21.5	20.6	19.7	18.7	191.0

INTERNAL RATE OF RETURN
(npv=0)

IRR	5.0

CASE: Sensitivity Case G

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Nat gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Escalation factors used for:
- manpower 1%
- maintenance 1%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.8%

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: GTM *
All figures in MM Canadian $ (4Q 1983)

	1986	1987	1988	1989	1990	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	TOTAL
Total Plant Cost	28.7	86.1	100.4	71.7																					
Working Capital				23.0																					
Total Investment	28.7	86.1	100.4	94.8																					
FEEDSTOCKS																									
Lignite																									
Wood																									
Peat																									
Natural Gas					116.9	149.4	152.8	156.3	159.8	163.5	167,2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
Electrolytic Hydrogen (Power Costs)																									
Total Feedstock Costs					116.9	149.4	152.8	156.3	159.8	163.5	167.2	171.0	174.8	178.8	182.9	187.0	191.3	195.6	200.1	204.6	209.2	214.0	218.8	223.8	
OPERATING COSTS																									
Manpower					5.5	5.6	5.7	5.8	6.0	6.1	6.2	6.3	6.4	6.6	6.7	6.8	7.0	7.1	7.3	7.4	7.6	7.7	7.9	8.0	
Maintenance					11.8	12.0	12.3	12.5	12.8	13.0	13.3	13.6	13.8	14.1	14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.8	17.2	
Utilities					1.0	1.3	1.3	1.3	1.4	1.4	1.4	1.4	1.4	1.4	1.4	1.5	1.5	1.5	1.5	1.5	1.5	1.5	1.6	1.6	
Taxes and Insurance					4.6	4.6	4.6	4.7	4.7	4.7	4.7	4.8	4.8	4.8	4.9	4.9	4.9	4.9	4.9	5.0	5.0	5.0	5.0	5.1	
Total Operating Costs					22.9	23.6	24.0	24.4	24.8	25.2	25.6	26.0	26.5	26.9	27.4	27.8	28.3	28.8	29.3	29.8	30.3	30.8	31.3	31.8	
CASHFLOW																									
Total Expenses					139.8	173.0	176.8	180.7	184.6	188.6	192.8	197.0	201.3	205.7	210.2	214.8	219.6	224.4	229.3	234.3	239.5	244.8	250.1	255.6	
Total Revenues					159.8	202.9	206.2	209.5	212.8	216.2	219.7	223.2	226.8	230.4	234.1	237.8	241.7	245.5	249.4	253.4	257.5	261.6	265.8	270.1	
Gross Profit					20.0	29.9	29.4	28.8	28.2	27.6	26.9	26.2	25.5	24.7	23.9	23.0	22.1	21.1	20.1	19.1	18.0	16.8	15.7	14.4	
Cash Flow	(28.7)	(86.1)	(100.4)	(94.8)	20.0	29.9	29.4	28.8	28.2	27.6	26.9	26.2	25.5	24.7	23.9	23.0	22.1	21.1	20.1	19.1	18.0	16.8	15.7	14.4	151.8

INTERNAL RATE OF RETURN
(npv=0)

IRR	4.0

CASE: Sensitivity Case H

Total plant cost: $287 MM
Working capital: $23MM
Total investment: $310 MM
Plant capacity: 2500 MTPD
On-stream days per year: 330
Plant utilization factor: 80% initial year
100% onward

Initial price of natural gas: 5.34 $/MSCF
Net gas yield: 0.0283 MSCF/liter
Initial selling price of methanol: 19.2 cents/liter

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- nat.gas price 2.27%
- methanol selling price 1.6%

FINANCIAL ANALYSIS

LIGNITE TO METHANOL

(Financial Analysis)

METHANOL ASSESSMENT PROGRAM

DISCOUNTED CASH FLOW ANALYSIS

Process: LTM Project (Onakawana Area)
All figures in MM Canadian $ (4Q 1983)

	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014
Total Plant Cost	87.7	263.1	307.0	219.3																				
Working Capital				12.0																				
Total Investment	87.7	263.1	307.0	231.3																				
FEEDSTOCKS																								
Lignite					39.6	47.7	53.3	53.5	53.8	54.1	54.4	54.6	54.9	55.2	55.4	55.7	56.0	56.3	56.6	56.8	57.1	57.4	57.7	58.0
Wood																								
Peat																								
Natural Gas																								
Electrolytic Hydrogen (Power Costs)																								
Total Feedstock Costs					39.6	47.7	53.3	53.5	53.8	54.1	54.4	54.6	54.9	55.2	55.4	55.7	56.0	56.3	56.6	56.8	57.1	57.4	57.7	58.0
OPERATING COSTS																								
Manpower					14.4	14.7	15.0	15.3	15.6	15.9	16.2	16.5	16.9	17.2	17.6	17.9	18.3	18.6	19.0	19.4	19.8	20.2	20.6	21.0
Maintenance					20.5	20.9	21.3	21.8	22.2	22.6	23.1	23.5	24.0	24.5	25.0	25.5	26.0	26.5	27.0	27.6	28.1	28.7	29.3	29.9
Utilities					3.0	3.6	4.1	4.1	4.2	4.2	4.2	4.3	4.3	4.4	4.4	4.5	4.5	4.6	4.6	4.6	4.7	4.7	4.8	4.8
Taxes and Insurance					6.2	6.2	6.3	6.3	6.3	6.4	6.4	6.4	6.5	6.5	6.5	6.5	6.6	6.6	6.6	6.7	6.7	6.7	6.8	6.8
Total Operating Costs					44.1	45.5	46.7	47.5	48.3	49.1	49.9	50.8	51.7	52.6	53.5	54.4	55.4	56.3	57.3	58.3	59.3	60.4	61.4	62.5
CASHFLOW																								
Total Expenses					83.7	93.2	99.9	101.0	102.1	103.2	104.3	105.4	106.6	107.7	108.9	110.1	111.4	112.6	113.9	115.1	116.4	117.8	119.1	120.5
Total Revenues					150.6	183.6	207.2	210.5	213.9	217.3	220.8	224.3	227.9	231.6	235.3	239.1	242.9	246.8	250.7	254.7	258.8	262.9	267.1	271.4
Gross Profit					66.9	90.4	107.3	109.5	111.8	114.2	116.5	118.9	121.4	123.8	126.4	128.9	131.5	134.2	136.8	139.6	142.3	145.2	148.0	150.9
Cash Flow	(87.7)	(263.1)	(307.0)	(231.3)	66.9	90.4	107.3	109.5	111.8	114.2	116.5	118.9	121.4	123.8	126.4	128.9	131.5	134.2	136.8	139.6	142.3	145.2	148.0	150.9
INTERNAL RATE OF RETURN (npv=0)																								
IRR	9.5																							

Total plant cost: $877 MM
Working capital: $12MM
Total investment: $889 MM
Plant capacity: 2500 MTPD
On-stream days per year: 320
Plant utilization factor: 75% initial year
90% second year
100% onward

Initial price of lignite: $16.6/MT
Lignite yield: 0.00315 T/liter
Initial selling price of methanol: 19.9 cents/liter
(includes shipping cost to Southern Ontario
and by-product credits)

Escalation factors used for:
- manpower 2%
- maintenance 2%
- utilities 1%
- tax and insurance .5%
- lignite price .5%
- methanol selling price 1.6%

```
DEBT AND CAPITAL DISBURSEMENTS IN NOMINAL DOLLARS   (Millions of Cdn Dollars)
=================================================

DEBT       YEAR   $REAL    CONV    $NOM
           1991       0    1.72       0
           1992     132    1.84     243
           1993     307    1.97     605
           1994     231    2.10     486

                                            INT  BL-INT
CAPITAL    1991      88    1.72     151
           1992     263    1.84     484       8       1
           1993     307    1.97     605      74      10
           1994     219    2.10     460     147      20
                                ------- ------- -------
                                   1700     229      31

EQUITY     1991      88    1.72     151
           1992     131    1.84     241

DEBT STRUCTURE
==============

YEAR    PRINCIPAL    TERM    RATE    1997    1998    1999    2000    2001    2002    2003    2004    2005    2006    2007    2008    2009    2010    2011    2012    2013    2014

1991            0
1992          243      10   13.50      24      24      24      24      24      24      24      24      24      24
1993          605      10   13.50      60      60      60      60      60      60      60      60      60      60
1994          486      10   13.50      49      49      49      49      49      49      49      49      49      49
         ---------                ------- ------- ------- ------- ------- ------- ------- ------- ------- ------- ------- ------- ------- ------- ------- ------- ------- -------
             1333                     133     133     133     133     133     133     133     133     133     133
         =========                ======= ======= ======= ======= ======= ======= ======= ======= ======= ======= ======= ======= ======= ======= ======= ======= ======= =======

INTEREST
========

             1991    1992    1993    1994    1995    1996    1997    1998    1999    2000    2001    2002    2003    2004    2005    2006    2007    2008    2009    2010    2011

                0       8      74     147     180     180     171     153     135     117      99      81      63      45      27       9       0       0       0       0       0

PRINCIPAL             243     847    1333    1333    1333    1200    1067     933     600     667     533     400     267     133       0       0       0       0       0       0
```

DEPRECIATION

	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014
FIXED ASSETS																				
BUILDING	588	559	531	504	479	455	432	411	390	371	352	334	318	302	287	272	259	246	234	222
EQUIPMENT	1372	1235	1111	1000	900	810	729	656	591	532	478	431	387	349	314	282	254	229	206	185
TOTAL	1960	1793	1642	1504	1379	1265	1161	1067	981	902	830	765	705	651	601	555	513	475	439	407
DEPRECIATION																				
BUILDING @5%	29	28	27	25	24	23	22	21	20	19	18	17	16	15	14	14	13	12	12	11
EQUIPMENT @10%	137	123	111	100	90	81	73	66	59	53	48	43	39	35	31	28	25	23	21	19
TOTAL	167	151	138	125	114	104	95	86	79	72	65	60	55	50	46	42	38	35	32	30

CAPITAL COST ALLOWANCE

UNDEP. CAPITAL COST

	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014
BUILDING	588	573	544	517	491	466	443	421	400	380	361	343	326	309	294	279	265	252	239	227
EQUIPMENT	1372	1029	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL	1960	1602	544	517	491	466	443	421	400	380	361	343	326	309	294	279	265	252	239	227

CAP.COST.ALLOW.

	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014
BUILDING @5%	15	29	27	26	25	23	22	21	20	19	18	17	16	15	15	14	13	13	12	11
EQUIPMENT (CL.29)	343	1029	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
TOTAL	358	1058	27	26	25	23	22	21	20	19	18	17	16	15	15	14	13	13	12	11

TIMING DIFFERENCES

	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014
TOTAL CCA	358	1058	27	26	25	23	22	21	20	19	18	17	16	15	15	14	13	13	12	11
TOTAL DEPRECIATION	167	151	138	125	114	104	95	86	79	72	65	60	55	50	46	42	38	35	32	30
TIMING DIFFERENCE	191	907	-110	-99	-89	-80	-72	-65	-59	-53	-47	-43	-38	-34	-31	-28	-25	-23	-20	-18
DEFER. TAX CR(DR)	86	408	-50	-45	-40	-36	-33	-29	-26	-24	-21	-19	-17	-16	-14	-13	-11	-10	-9	-8

OJECT E STA

FOR THE YEARS 1991-2014

	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014
REVENUES:																								
SALES					340	443	534	579	632	685	747	810	882	960	1041	1133	1232	1340	1458	1584	1722	1871	2032	2207
	0	0	0	0	340	443	534	579	632	685	747	810	882	960	1041	1133	1232	1340	1458	1584	1722	1871	2032	2207
FEEDSTOCK COSTS:																								
LIGNITE					90	116	137	146	159	171	182	199	213	228	244	265	284	304	331	354	379	405	441	472
	0	0	0	0	90	116	137	146	159	171	182	199	213	228	244	265	284	304	331	354	379	405	441	472
OPERATING COSTS:																								
MANPOWER					32	36	39	41	47	51	54	61	66	70	80	85	91	103	110	118	133	142	160	171
MAINTENANCE					47	51	54	61	65	73	78	87	93	99	111	123	132	147	157	174	186	206	221	244
UTILITIES					7	10	10	11	12	13	14	14	15	17	18	19	22	25	27	31	33	36	38	41
TAX/INSURANCE					14	14	15	17	18	19	20	22	27	29	31	33	36	38	41	43	47	50	53	57
DEPRECIATION					167	151	138	125	114	104	95	86	79	72	65	60	55	50	46	42	38	35	32	30
INTEREST					220	225	222	212	196	175	145	106	69	45	27	9	0	0	0	0	0	0	0	0
	0	0	0	0	485	487	478	467	452	433	405	377	349	332	332	329	336	363	380	408	437	469	504	543
INCOME BEFORE TAX	0	0	0	0	-235	-160	-81	-33	20	81	159	234	320	401	466	538	613	674	746	822	906	996	1087	1192
ADD: DEPRECIATION					167	151	138	125	114	104	95	86	79	72	65	60	55	50	46	42	38	35	32	30
LESS: CCA					358	1058	27	26	25	23	22	21	20	19	18	17	16	15	15	14	13	13	12	11
TAXABLE INCOME	0	0	0	0	-427	-1066	30	66	110	162	231	299	379	454	513	581	651	708	777	850	931	1019	1107	1210
TAX RECOVERY	0	0	0	0	192	480	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
INCOME TAX-CURRENT	0	0	0	0	0	0	13	30	49	73	104	135	170	204	231	261	293	319	350	382	419	459	498	545
-DEFERRED					86	408	-50	-45	-40	-36	-33	-29	-26	-24	-21	-19	-17	-16	-14	-13	-11	-10	-9	-8
	0	0	0	0	-106	-72	-36	-15	9	37	72	105	144	180	210	242	276	303	336	370	408	448	489	536
NET INCOME	0	0	0	0	-130	-88	-45	-18	11	45	87	129	176	220	256	296	337	371	410	452	498	548	598	656

PROJECTED CASH FLOWS FOR THE YEARS 1991-2014

	1991	1992	1993	1994	1995	1996	1997	1998	1999	2000	2001	2002	2003	2004	2005	2006	2007	2008	2009	2010	2011	2012	2013	2014
OPERATIONS																								
CASH RECEIPTS:																								
REVENUES	0	0	0	0	340	443	534	579	632	685	747	810	882	960	1041	1133	1232	1340	1458	1584	1722	1871	2032	2207
DISBURSEMENTS:																								
LIGNITE	0	0	0	0	90	116	137	146	159	171	182	199	213	228	244	265	284	304	331	354	379	405	441	472
MANPOWER	0	0	0	0	32	36	39	41	47	51	54	61	66	70	80	85	91	103	110	118	133	142	160	171
MAINTENANCE	0	0	0	0	47	51	54	61	65	73	78	87	93	99	111	123	132	147	157	174	186	206	221	244
UTILITIES	0	0	0	0	7	10	10	11	12	13	14	14	15	17	18	19	22	25	27	31	33	36	38	41
TAX/INSURANCE	0	0	0	0	14	14	15	17	18	19	20	22	27	29	31	33	36	38	41	43	47	50	53	57
INTEREST	0	9	84	167	220	225	222	212	196	175	145	106	69	45	27	9	0	0	0	0	0	0	0	0
TAXES	0	0	0	0	0	0	0	0	0	0	0	0	0	204	231	261	293	319	350	382	419	459	498	545
	0	9	84	167	409	451	477	488	497	500	493	490	483	692	741	796	858	935	1015	1103	1197	1298	1412	1530
EXCESS OF RECEIPTS OVER DISBURSEMENTS FROM OPERATIONS	0	-9	-84	-167	-69	-8	57	92	134	185	254	320	399	268	300	336	374	405	442	481	525	573	621	677
FUNDING																								
RECEIPTS:																								
EQUITY	151	241	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DEBT	0	243	605	486	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
BANK LINE	0	9	93	260	329	337	414	455	454	402	282	95												
	151	493	698	746	329	337	414	455	454	402	282	95	0	0	0	0	0	0	0	0	0	0	0	0
DISBURSEMENTS:																								
CAPITAL	151	484	605	460	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0	0
DEBT REPAYMENT							133	133	133	133	133	133	133	133	133	133	0	0	0	0	0	0	0	0
BANK LINE			9	93	260	329	337	414	455	454	402	282	95	0	0	0	0	0	0	0	0	0	0	0
	151	484	614	553	260	329	470	547	588	587	535	415	228	133	133	133	0	0	0	0	0	0	0	0
EXCESS OF RECEIPTS OVER DISBURSEMENTS FOR FUNDING	0	9	84	193	69	8	-56	-92	-134	-185	-253	-320	-228	-133	-133	-133	0	0	0	0	0	0	0	0
EXCESS-RECEIPTS OVER DISBURSEMENTS	0	0	0	26	0	0	0	-1	0	0	0	0	170	135	167	203	374	405	442	481	525	573	621	677
OPENING CASH	0	0	0	0	26	26	26	26	26	26	26	26	26	196	331	498	701	1076	1481	1923	2404	2929	3502	4123
CLOSING CASH	0	0	0	26	26	26	26	26	26	26	26	26	196	331	498	701	1076	1481	1923	2404	2929	3502	4123	4800